U.S. DEPARTMENT OF COMMERCE
National Technical Information Service

PB-259 150

Process Design Manual for Phosphorus Removal

Shimek, Roming, Jacobs and Finklea, Dallas, Tex

Prepared for

Environmental Protection Agency, Cincinnati, Ohio Office of Technology Transfer

Apr 76

300155

EPA 625/1-76-001a

PB 259 150

PROCESS DESIGN MANUAL
FOR
PHOSPHORUS REMOVAL

U.S. ENVIRONMENTAL PROTECTION AGENCY
Technology Transfer

April 1976

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

288

ACKNOWLEDGMENTS

The original edition of this Design Manual (October 1971) was prepared for Technology Transfer by Black & Veatch, Consulting Engineers. This first revision to the basic text was prepared for Technology Transfer by the firm of Shimek, Roming, Jacobs and Finklea. Major EPA contributors were E. F. Barth, J. M. Smith, C. A. Brunner, and J. B. Farrell of the U.S. EPA Office of Research and Development, Cincinnati, Ohio. Major reviewers were D. J. Lussier, J. C. Dyer, and R. M. Madancy of Technology Transfer.

Revisions to Chapters 4, 5, 10 and 11 have been made by J. Laughlin of Shimek, Roming, Jacobs and Finklea in order to update the information in these chapters. Appendix A has been added to describe an EPA computer model for evaluating phosphorus removal strategies.

NOTICE

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BIBLIOGRAPHIC DATA SHEET

1. Report No.
EPA/625/1-76/001a

2.

3. Recipient's Accession No.
PB-259 150

4. Title and Subtitle
Process Design Manual for Phosphorus Removal.

5. Report Date
April 1976

6.

7. Author(s)

8. Performing Organization Rept. No.

9. Performing Organization Name and Address
Shimek, Roming, Jacobs and Finklea
Dallas, Texas

10. Project/Task/Work Unit No.

11. Contract/Grant No.

12. Sponsoring Organization Name and Address

13. Type of Report & Period Covered

14.

15. Supplementary Notes
See also PB-214 553. (PC A13/MF A01).

16. Abstracts

The manual discusses phosphorus removal methods that have been found effective and practical for use at treatment plants. All the methods included involve chemical precipitation of the phosphorus and removal of the resultant precipitate. Precipitants include salts of aluminum, iron, and lime. The practical points of addition are before the primary settler, in the aerator of an activated sludge plant, before the final settler, or in a tertiary process. Included in the discussion of each treatment method is a description of the method, pilot or full-scale performance data, equipment requirements, design parameters and costs.

17. Key Words and Document Analysis. 17a. Descriptors

*Sewage treatment
*Phosphorus
Chemical removal(Sewage treatment)
Manuals
Design
Technology transfer
Chemical removal(Sewage treatment)
Precipitation(Chemistry)
Efficiency
Performance
Cost analysis
Pilot plants
Waste water
Municipalities
Aluminum
Iron
Calcium oxides
Equipment
Activated sludge process

17b. Identifiers/Open-Ended Terms

*Sewage treatment plants
*Tertiary treatment

PRICES SUBJECT TO CHANGE

17c. COSATI Field/Group 13B

ii-A

18. Availability Statement
National Technical Information Service
Springfield, VA 22161

19. Security Class (This Report)
UNCLASSIFIED

20. Security Class (This Page)
UNCLASSIFIED

21. No. of Pages
288

22. Price PC$9.25 /MF$3.00

FORM NTIS-35 (REV. 10-73) ENDORSED BY ANSI AND UNESCO. THIS FORM MAY BE REPRODUCED USCOMM-DC 8265-P74

ABSTRACT

The discharge of phosphorus-containing wastewaters into the surface waters of the United States has contributed to their over-fertilization and eutrophication. As a result, efforts are now being made to remove phosphorus from wastewater.

This manual discusses phosphorus removal methods that have been found effective and practical for use at treatment plants. All the methods included involve chemical precipitation of the phosphorus and removal of the resultant precipitate. Precipitants include salts of aluminum and iron, and lime. The practical points of addition are before the primary settler, in the aerator of an activated sludge plant, before the final settler, or in a tertiary process.

Included in the discussion of each treatment method is a description of the method, pilot or full-scale performance data, equipment requirements, design parameters, and costs. This information should be of value to designers, municipal officials, regulatory agencies, city planners, and treatment plant operators.

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FOREWORD

The formation of the United States Environmental Protection Agency marked a new era of environmental awareness in America. This Agency's goals are national in scope and encompass broad responsibility in the area of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest available techniques, will be adequate to meet the future water quality objectives and to ensure continued protection of the nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry a new source of information to be used in the planning, design and operation of present and future wastewater treatment facilities. It is recognized that there are a number of design manuals, manuals of standard practice, and design guidelines currently available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of several available through the Technology Transfer Office of EPA to describe recent technological advances and new information. This particular manual was initially issued in October of 1971 and this edition represents the first revision to the basic text. Future editions will be issued as warranted by advancing state-of-the-art to include new data as it becomes available, and to revise design criteria as additional full-scale operational information is generated.

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Chapter 1

INTRODUCTION

1.1 Purpose

The technology for removal of phosphorus from wastewater has developed rapidly in the last few years. The need for practical phosphorus removal procedures is a result of the over-fertilization and eutrophication of the Country's surface waters. This further resulted in the establishment of state water quality standards that limit the concentration of phosphorus in receiving waters. As an example, states that have streams tributary to the Great Lakes have established wastewater effluent phosphorus limits or require a fixed percentage reduction of phosphorus. Several other states have also established phosphorus standards.

The Environmental Protection Agency has sponsored many research and demonstration studies at several cities in the past few years to advance the knowledge of phosphorus removal. Local and state governments and private industries have also contributed to this work. This manual is intended to summarize process design information for the best developed removal methods that have resulted from this governmental and private effort.

1.2 Scope

This manual discusses a number of phosphorus removal methods that involve chemical precipitation. Phosphorus removal obtainable by biological activity alone is not included. Treatment methods in which phosphorus removal occurs, but is not a principal objective, are also omitted. The latter group of processes includes ion exchange, reverse osmosis and other demineralization treatments which at present are more closely associated with wastewater renovation and reuse than with pollution control. These will be included in updated versions of the manual when appropriate. One application of a precipitation method that has been omitted and which is of growing interest is complete lime treatment of raw sewage. This application will be included when the manual is updated.

The information included on chemical methods of phosphorus removal was obtained from the available literature, progress reports of demonstration studies, and private communications with investigators actively working in the field. Design guidelines have been developed from these sources.

The information contained in this manual is oriented toward design methods and operating procedures for removal of phosphorus from wastewater to comply with state water quality standards. It is recognized that present state standards may eventually be superseded with more stringent requirements for phosphorus removal. Such revisions in state standards must be considered in design of treatment facilities to minimize future modifications to treatment plant structures.

Engineers are often faced with circumstances that require the design of treatment facilities with a limited amount of data, testing, and time. Such circumstances require a more conservative design to allow for contingencies. Since most phosphorus removal methods are of recent development, they fall into this category. Conservative values are given, therefore, when stating design parameters.

Chapter 2

BASIC DESIGN CONSIDERATIONS

2.1 Sources of Phosphorus

Domestic wastewater normally has a substantial concentration of phosphorus with the primary sources being a result of man's activities in the home. Human wastes such as feces, urine, and waste food disposal account for about 30 to 50% of the phosphorus in domestic wastewater (1). Detergents containing phosphate builders and used principally for laundering of clothes account for the remainder of phosphorus, or about 50 to 70%. Other sources of phosphorus may cause deviation from these percentages. For example, where sodium hexametaphosphate or other phosphorus compounds are used as corrosion and scale control chemicals in water supplies, the phosphorus added will be present in the same concentration, although not necessarily in the same form, as in the water supply. This source can account for 2 to 20% of the total phosphorus present in the wastewater.

Industrial wastewater discharges to the sanitary sewer system may either dilute or increase the concentration of phosphorus present in the combined wastewater. For example, wastewaters from pulp and paper mills may be phosphorus deficient and therefore dilution will occur. On the other hand, wastewaters discharged from some industries, such as potato processing plants, may contain high concentrations of phosphorus and consequently may increase the phosphorus concentration of the combined wastewater.

The quantity of phosphorus resulting from human excretions reportedly ranges from 0.5 to 2.3 lb per capita per year (2). The mean annual excretion is estimated to be about 1.2 lb per capita. The mean annual contribution of phosphorus from synthetic detergents with phosphate builders is estimated to be about 2.3 lb per capita, at present. Thus, exclusive of industrial wastes and other phosphorus sources, such as water softening or sequestering agents, the domestic phosphorus contribution to wastewater is about 3.5 lb per capita per year.

On this basis, the wastewater phosphorus concentration may be estimated for a community where there are no phosphorus data available. Assuming a population of 5,000, the total phosphorus discharged equals 5,000 x 3.5 = 17,500 lb/year, or 48 lb/day. If a wastewater flow of 100 gal. per capita per day is assumed, the total flow is 0.5 million gal. per day (mgd) and the phosphorus concentration in the wastewater is as follows:

$$\frac{48}{0.5 \times 8.34} = 11.5 \text{ mg/l (as P)}$$

The average total phosphorus concentration in domestic raw wastewater is found to be about 10 mg/l expressed as elemental phosphorus (P).

The above information is provided to serve as a rough guide to engineers and not as a basis of design. Sampling of the wastewater and analyses for phosphorus are recommended in all cases. It may also be necessary to survey the industries that will be discharging wastewater to the sanitary sewer system, to determine their influence on the concentration of phosphorus in the combined wastewater.

Among industrial sources of phosphorus may be potato processing wastes (3), fertilizer manufacturing wastes (4), animal feedlot wastes (5), certain metal finishing wastes (1, 6), flour processing wastes (7), dairy wastes (8, 9), commerical laundry wastes (10), and slaughterhouse wastes (11). The amount of phosphorus may be quite variable depending on the specific industrial plant.

2.2 State Regulatory Agencies

Each state has at least one organization responsible for water pollution prevention, control, and abatement activities. In about half of the 50 states this responsibility is divided between two state agencies. Usually one of these is the state health department and the other an organization charged specifically by statute with the conduct of the state's water pollution control program. Although there has been an intensification of Federal activities in water pollution control over the past twenty years, much of the responsibility for program operations rests with the states. Invariably the appropriate state agencies should be consulted in connection with the planning and conduct of water pollution control programs at cities and other communities. The Environmental Protection Agency regional office personnel work closely with state agencies and can give advice regarding the proper state agency contacts.

2.3 State Standards

As of June 1971, sixteen states had adopted wastewater effluent phosphorus standards. In general, these standards have taken the form of an effluent concentration limit or a requirement for a specified percentage reduction in the phosphorus concentration in the raw wastewater. In most cases, effluent concentration limits range from 0.1 to 2.0 mg/l as P, with many established at 1.0 mg/l. Percentage reduction requirements range from 80 to 95%.

It should be noted that neither an effluent nor a percentage-reduction standard actually limits the phosphorus load in terms of pounds of phosphorus discharged per day. Load (lb/day) is a direct function of both effluent P concentration and daily effluent volume. Assuming the effluent concentration standard is being met, an increase in the daily flow of effluent will produce a proportionate increase in the mass of phosphorus discharged daily to the receiving waters. If the phosphorus load then becomes intolerable, adjustment downward of the effluent concentration standard will be necessary. Similarly, adjustment upward of a percentage-reduction standard will be required if the phosphorus load becomes objectionable as a result of increased wastewater flow or increased P concentration in the raw wastewater or both.

Effluent concentration standards can be derived from receiving water standards as indicated in the following example.

Receiving Water	
Phosphorus concentration upstream	0.01 mg/l

Receiving Water - (continued)	
Maximum allowable phosphorus concentration	0.05 mg/l
Low flow	140 ft^3/sec (90 mgd)

Wastewater	
Average daily flow	4.8 mgd
Effluent concentration standard	X

Mass Balance

$$(0.01)(8.34)(90) + (X)(8.34)(4.8) = (0.05)(8.34)(90+4.8)$$

$$X = 0.8 \text{ mg/l P}$$

Implicit in the above calculation is the assumption that the effluent is mixed thoroughly with the waters of the receiving stream. The low flow selected is not necessarily the low flow of record. It is usually the low flow averaged over a period of 10 days or less and having a return period of once in 10 years.

2.4 Provision for Future Plant Expansion and More Stringent Effluent Requirements

The plant site is an important consideration in the design of treatment works. It is highly desirable that sufficient site area be obtained to permit initial construction of the required system without crowding and to provide space for future expansion. Treatment systems are likely to become larger and more complex in the future rather than less so, and failure to obtain ample site area initially is almost certain to create costly problems later.

Reduction of the allowable effluent phosphorus concentration may be necessary in some states as increases in wastewater flow or phosphorus concentration of the raw wastewater occur. In some instances, merely increasing the chemical dosage may be the only necessary change. On the other hand, it may be necessary to add an additional treatment sequence such as filtration to the existing facilities. In nearly every case, planning for plant expansion or more stringent effluent requirements will result in economy. Such planning should include selection of a plant layout that allows space for addition of facilities without removing existing processes from service.

Other provisions that should be considered are space within chemical buildings for additional coagulant storage and feeding equipment, space for polymer storage and feed equipment, and adequate space for handling the additional solids generated by phosphorus removal processes.

2.5 Flow Equalization

The cyclic nature of wastewater flows, in terms of volume and strength, is well established. While the concept of flow equalization has been employed in the field of water supply, and in the treatment of some industrial wastes, it has not been widely accepted in the municipal pollution control field. Anticipated problems with solids settling, odor, and septicity can be cited as the major factors limiting its use.

The advent of stricter stream standards, which sometimes require removal of contaminants during peak flows, the elimination of plant bypassing, and the increased removal efficiencies that are possible when biological or chemical treatment processes are operated at near steady-state conditions, have all favored the increased use of flow-smoothing, and flow equalization devices ahead of or integral with the design of major treatment works.

There are two major objectives in the design of flow equalization basins. The first of these is simply to dampen the diurnal flow variation that normally exists in typical municipal wastewater collection systems to achieve a constant or nearly constant flow rate through the downstream treatment processes. In this type of system, little consideration is given to controlling the concentration changes that take place during storage. The major design factors are supplying enough air to keep the basin aerobic and to provide adequate mixing to prevent solids deposition.

The second objective of flow equalization is to provide the capacity to distribute shock loads of toxic or process inhibiting substances over a reasonable period of time so as to prevent system failure and to minimize the periodic discharge of harmful contaminants to the receiving stream or surface impoundment. Time-dependent concentration profiles and flow-through curves are normally used to analyze the flow characteristics of these systems and to determine the effect of tank geometry, placement of effluent weirs and mixing regime on changes in contaminant concentrations through the basin.

In all cases the added costs of flow equalization must be measured against the possible reduction in downstream process costs and the increased efficiencies that can be achieved by operating these processes under constant loading conditions.

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Chapter 3

THEORY OF PHOSPHORUS REMOVAL BY CHEMICAL PRECIPITATION

3.1 Forms and Measurement of Phosphorus

Phosphorus is found in wastewater in three principal forms: orthophosphate ion, polyphosphates or condensed phosphates, and organic phosphorus compounds. Orthophosphate was in the past often considered to be PO_4^{3-} and was reported in this way. Actually, there are a number of forms of orthophosphate in equilibrium, with the predominant form changing as pH changes. At the usual pH of municipal wastewater, the predominant form is HPO_4^{2-}. With the present practice of reporting phosphorus as P, one does not usually have to be concerned with the actual form. It is well to remember, however, that the predominant form does change with pH. The removal of phosphorus with lime results primarily from pH increase. Polyphosphates can be looked upon as polymers of phosphoric acid from which water has been removed. Materials of various molecular weight are in widespread use. Complete hydrolysis results in formation of orthophosphate. The chemistry of organic phosphorus compounds is complicated. Their decomposition also leads to orthophosphate.

In raw sewage there are substantial amounts of all three principal phosphorus forms. During biological treatment, significant changes take place. As organic materials are decomposed, their phosphorus content is converted to orthophosphate. On the other hand, inorganic phosphates are utilized in forming biological floc. The polyphosphates are for the most part converted to orthophosphate. The result is that in a well treated secondary effluent, a large fraction of the phosphorus is present as orthophosphate. From the standpoint of removal by precipitation, this is fortunate since the orthophosphate is the easiest form to precipitate.

Because phosphorus will be present in a variety of forms, both organic and inorganic, the only satisfactory measure of treatment plant removal efficiency must be based on total phosphorus entering the plant in the raw wastewater and total phosphorus discharged in the plant effluent. Inasmuch as all colorimetric tests for phosphorus depend upon the formation of an orthophosphate color complex, it is essential in the determination of total phosphorus, to convert any polyphosphates (condensed phosphates) and organic phosphorus compounds present in the sample to the orthophosphate form. This conversion is accomplished by acid digestion of the sample in the presence of a strong oxidizing agent. *Standard Methods*, 13th edition (1), describes several procedures for determining total phosphorus as well as the various forms of phosphorus in wastewater. A suitable procedure for measuring total phosphorus, including the preliminary digestion stage, may be selected from this reference. An automated method is described in *FWPCA Methods for Chemical Analysis of Water and Wastes (2)*.

3.2 Chemistry of Removal by Precipitation

The materials found practical for phosphorus precipitation include the ionic forms of aluminum, iron, and calcium. The calcium is added as lime and the hydroxyl ions also have a role in phosphorus removal. The chemistry of phosphorus removal by interaction with metallic ions is complex. In the interests of brevity and simplicity, it is assumed in the following discussion that phosphorus reacts as the orthophosphate form, PO_4^{3-}, and that the reaction products are as indicated. The polyphosphates and organic phosphorus are also removed, probably by a combination of more complex reactions and sorption on floc particles. The reactions presented and the associated computations are principally for illustrative purposes and do not purport to represent the true reactants, products and reaction mechanisms. Calculated requirements for phosphorus precipitants must be viewed as no more than crude estimates subject to large variations in actual practice because the forms of phosphorus and the precipitates may be substantially different from those assumed.

3.2.1 Aluminum Compounds as Phosphorus Precipitants

Aluminum ions can combine with phosphate ions to form aluminum phosphate as follows:

$$Al^{3+} + PO_4^{3-} \longrightarrow \underline{AlPO_4}$$

The above equation indicates that the mole ratio for $Al:PO_4$ is 1:1. Inasmuch as the mole ratio of $P:PO_4$ is also 1:1, the mole ratio for Al:P is 1:1 or $\frac{Al}{P} = 1$ when both aluminum and phosphorus are expressed in terms of gram-moles or lb-moles. On a weight, rather than a mole basis, this means that 27 lb of Al will react with 95 lb of PO_4 to form 122 lb of $AlPO_4$. Since each 95 lb (1 lb-mole) of PO_4 contains 31 lb (1 lb-mole) of P, the weight relationship between Al and P is 27 lb of Al to 31 lb of P or 0.87 for this reaction.

The principal source of aluminum for use in phosphorus precipitation is "alum", a hydrated aluminum sulfate, having the approximate formula $Al_2(SO_4)_3 \cdot 14H_2O$ (molecular weight of 594). The chemical, which is also known as "filter alum" or "papermaker's alum", averages about 17% soluble aluminum expressed as Al_2O_3 or about 9.1% expressed as Al. Its reaction with PO_4^{3-} may be written as follows:

$$Al_2(SO_4)_3 \cdot 14H_2O + 2PO_4^{3-} \longrightarrow \underline{2AlPO_4} + 3SO_4^{2-} + 14H_2O$$

The sulfate remains in solution as SO_4^{2-}. The above reaction indicates that 1 lb-mole of alum (594 lb) will react with 2 lb-moles (190 lb) of PO_4^{3-} containing 62 lb phosphorus to form 2 lb-moles (244 lb) of $AlPO_4$. The weight ratio of alum to phosphorus is, therefore, 9.6:1. The alum requirement per pound of phosphorus may also be derived from the Al:P mole ratio as follows:

Mole ratio Al:P = 1:1
Therefore weight ratio Al:P = 27:31 = 0.87:1
Alum contains 9.1% Al
Therefore, alum required per lb of P = $\frac{0.87}{0.091}$ = 9.6 lb

Stumm and Morgan (3), state that the solubility of $AlPO_4$ is pH dependent and varies as follows:

pH	Approximate Solubility (mg/l)
5	0.03
6	0.01 (minimum solubility)
7	0.3

The optimum pH for removal of phosphorus probably lies in the range of 5.5 to 6.5, although some removal occurs above pH 6.5.

Addition of alum will lower the pH of wastewater because of neutralization of alkalinity and release of carbon dioxide. The extent of pH reduction will depend principally on the alkalinity of the wastewater. The higher the alkalinity, the less is the reduction in pH for a given alum dosage. Most wastewaters contain sufficient alkalinity so that even large alum dosages will not lower the pH below about 6.0 to 6.5. In exceptional cases of low wastewater alkalinity, pH reduction may be so great that addition of an alkaline substance, such as sodium hydroxide, soda ash, or lime will be required.

Adjustment of the pH downward can be accomplished by the addition of sulfuric acid but this complicates the treatment process and it may be preferable simply to use a higher alum dosage.

Bench, pilot, and full-scale studies have shown that considerably higher than stoichiometric quantities of alum usually are necessary to meet phosphorus removal objectives. A competing reaction, responsible for the pH reduction mentioned above, at least partially accounts for the excess alum requirement. It occurs as follows:

$$Al_2(SO_4)_3 \cdot 14H_2O + 6HCO_3^- \longrightarrow \underline{2Al(OH)_3} + 6CO_2 + 14H_2O + 3SO_4^{2-}$$

The following ratios of alum (9.1%Al) to phosphorus are believed reasonably representative for alum treatment of municipal wastewater (4).

P Reduction Required	Al:P Mole Ratio	Al:P Weight Ratio	Alum:P Weight Ratio
75%	1.38:1	1.2:1	13:1
85%	1.72:1	1.5:1	16:1
95%	2.3:1	2.0:1	22:1

To achieve 85% P removal from a wastewater containing 11 mg/l of P the alum dosage needed would be (16)(11) = 176 mg/l or 1470 lb/10^6 gal.

Sodium aluminate will also serve as a source of aluminum for the precipitation of phosphorus. The chemical formula for sodium aluminate is $Na_2Al_2O_4$ or $NaAlO_2$.

One commercial form is the granular trihydrate, which may be written $Na_2O\ Al_2O_3 \cdot 3H_2O$ and which contains about 46% Al_2O_3 or 24% Al.

The sodium aluminate - phosphate reaction may be represented as follows:

$$Na_2O \cdot Al_2O_3 + 2PO_4^{3-} + 4H_2O \longrightarrow \underline{2AlPO_4} + 2NaOH + 6OH^-$$

In contrast to alum, which reduces pH, a rise in pH may be expected upon addition of sodium aluminate to wastewater.

The above reaction indicates an Al:P mole ratio of 1:1, an Al:P weight ratio of 0.87:1, and a sodium aluminate to P weight ratio of about 3.6:1.

As with the use of alum, mole and weight ratios somewhat in excess of the theoretical must be anticipated in practice.

3.2.2 Iron Compounds as Phosphorus Precipitants

Both ferrous (Fe^{2+}) and ferric (Fe^{3+}) ions can be used in the precipitation of phosphorus. With Fe^{3+} a reaction can be written similar to that shown earlier for precipitation of aluminum phosphate. A 1:1 mole ratio of $Fe:PO_4$ results. Since the molecular weight of iron is 55.85, the weight ratio of Fe:P is 1.8:1. Just as in the case of aluminum, a larger amount of iron is required in actual situations than the chemistry of the reaction predicts. With Fe^{2+} the situation is more complicated and not fully understood. Ferrous phosphate can be formed. The mole ratio of $Fe:PO_4$ would be 3:2. Experimental results indicate, however, that when Fe^{2+} is used, the mole ratio of Fe:P will be essentially the same as when Fe^{3+} is used.

A number of iron salts are available for use in phosphorus precipitation. These include ferrous sulfate, ferric sulfate, ferric chloride, and pickle liquor. Additional discussion of these is given later. All will lower the pH of wastewater because of neutralization of alkalinity. Pickle liquor contains substantial amounts of free sulfuric acid or hydrochloric acid.

Iron salts are most effective for phosphorus removal at certain pH values. For Fe^{3+} the optimum pH range is 4.5 to 5.0. This is an unrealistically low pH, not attained in most wastewaters. Significant removal of phosphorus can be attained at higher pH. For Fe^{2+} the optimum pH is about 8 and good phosphorus removal can be obtained between 7 and 8. The acidity of ferrous salts, and especially pickle liquor, necessitates addition of lime or sodium hydroxide for good results. Where the water is aerated following Fe^{2+} addition, the use of a base may not be necessary.

3.2.3 Lime as a Phosphorus Precipitant

Calcium ion reacts with phosphate ion in the presence of hydroxyl ion to form hydroxyapatite. This material has a variable composition, but an approximate equation for its formation can be written as follows, assuming in this case that the phosphate present is HPO_4^{2-}:

$$3\ HPO_4^{2-} + 5Ca^{2+} + 4OH^- \longrightarrow \underline{Ca_5(OH)(PO_4)_3} + 3\ H_2O$$

The reaction is pH dependent. The solubility of hydroxyapatite is so low, however, that even at a pH as low as 9.0, a large fraction of the phosphorus can be removed. In lime treatment of wastewater, the operating pH may be predicated on the ability to obtain good suspended solids removal rather than on phosphorus removal.

Although it is possible to calculate an approximate lime dose for phosphorus removal, this is generally not necessary. In contrast to iron and aluminum salts, the lime dose is largely determined by other reactions that take place when the pH of wastewater is raised. Some of these reactions are discussed later. Only in waters of very low bicarbonate alkalinity would the phosphate precipitation reaction consume a large fraction of the lime added.

3.3 References

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Chapter 4

PHOSPHORUS REMOVAL BY MINERAL ADDITION BEFORE THE PRIMARY SETTLER

4.1 General Considerations

Removal of phosphorus from wastewater can be accomplished by modifying the conventional primary treatment process to include chemical precipitation with aluminum or iron salts (1,2). The advantages of phosphorus removal during primary treatment include flexibility in chemical feeding, adequate reaction times and mixing conditions, flocculation and removal of more suspended solids and suspended BOD in the primary settler, and reduced loading of suspended solids and BOD to secondary treatment processes. Significant increases in removals of both suspended solids and BOD can occur concurrently with phosphorus removal. Phosphorus removal by chemical addition is amenable to automatic control using process instrumentation to measure waste flow and/or phosphorus concentration.

The disadvantage of addition of metal salts during primary treatment alone is that a significant amount of the phosphorus may not be in the ortho form and may not be easily precipitated. Higher phosphorus removals can be obtained with chemical addition beyond primary treatment and with chemical additions at several points.

The available precipitation-flocculation processes (3,4,5,6,7,8) include addition of ferrous or ferric chloride, ferrous or ferric sulfate, aluminum sulfate, or sodium aluminate, followed by flocculation using an anionic polymer to enhance solids separation. A strong base is often added between additions of ferrous iron and polymer to counteract the depression of pH by the ferrous compounds.

In order to provide effective chemical additions to primary treatment units, the following sequence is recommended:

1. Salt addition to raw sewage with thorough mixing. If a base is to be used, it should be added at least 10 seconds later.
2. Reaction for at least five minutes.
3. Addition of anionic organic polymer followed by flash mixing for 20 to 60 seconds.
4. Mechanical or air flocculation for 1 to 5 minutes.
5. Gentle delivery of flocculated wastewater to sedimentation units.

Mixing may be carried out in specially designed basins or advantage may be taken of existing high turbulence areas such as in a Parshall flume. Flocculation may similarly be carried out in specially constructed equipment or may be allowed to take place within some part of an existing system. Schematic diagrams of 2 plants, which demonstrate how existing equipment can be used with mineral addition, are shown in Figures 4-1 and 4-2.

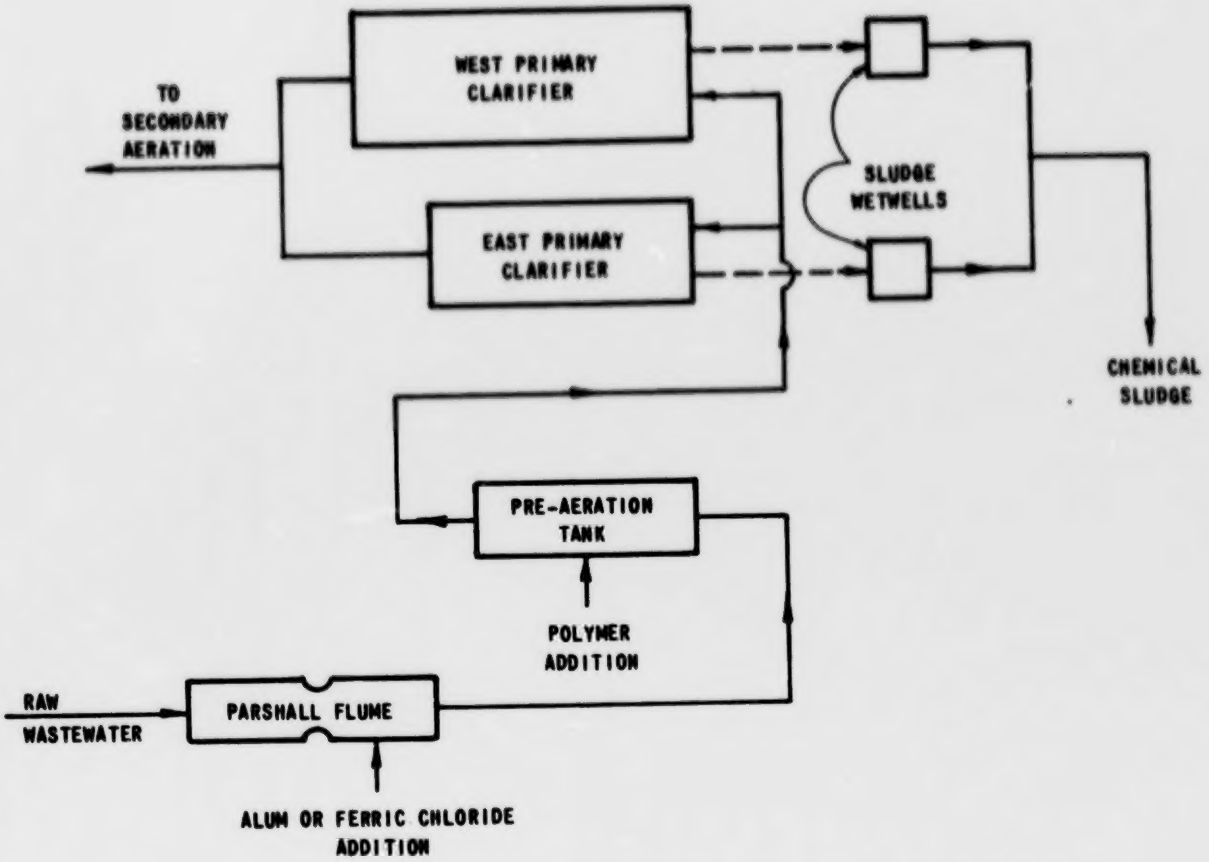

WEST PRIMARY CLARIFIER
EAST PRIMARY CLARIFIER
TO SECONDARY AERATION
SLUDGE WETWELLS
CHEMICAL SLUDGE
PRE-AERATION TANK
POLYMER ADDITION
RAW WASTEWATER
PARSHALL FLUME
ALUM OR FERRIC CHLORIDE ADDITION

FIGURE 4-1 LEBANON, OHIO PRIMARY TREATMENT AND CHEMICAL FEEDING SCHEMATIC

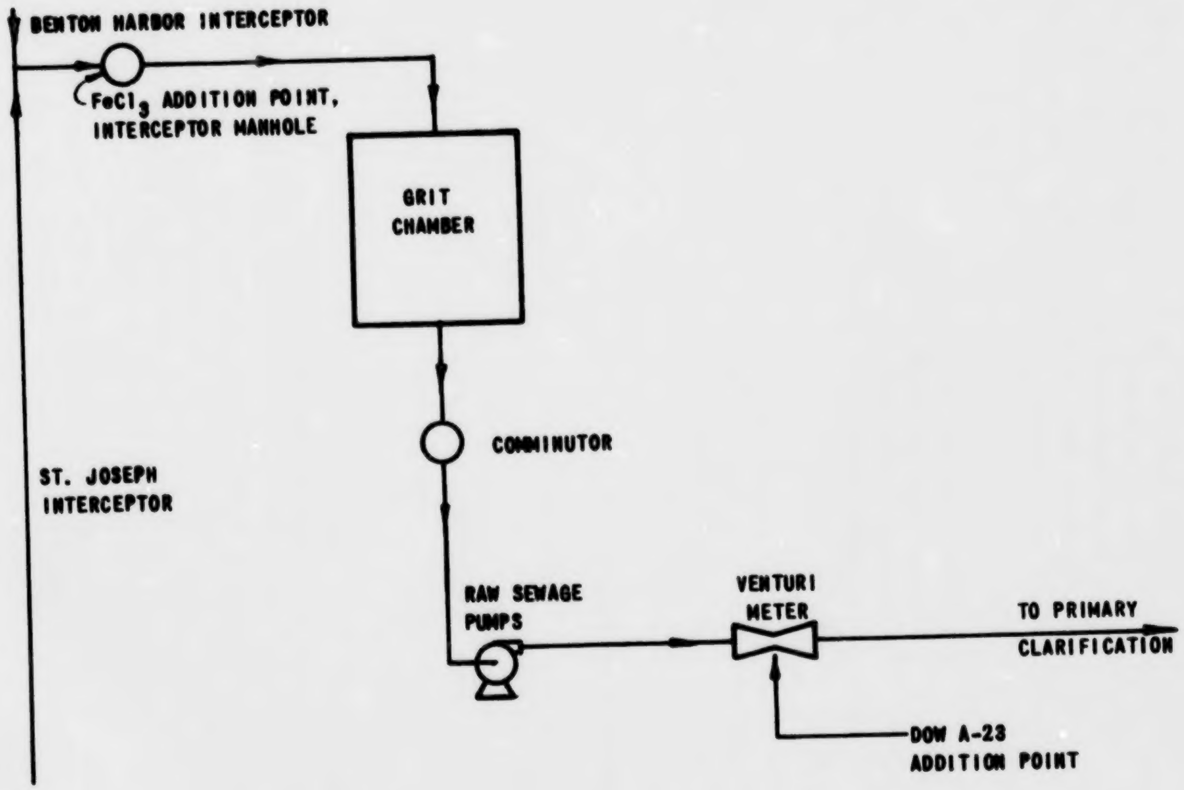

BENTON HARBOR INTERCEPTOR
FeCl3 ADDITION POINT, INTERCEPTOR MANHOLE
GRIT CHAMBER
COMMINUTOR
ST. JOSEPH INTERCEPTOR
RAW SEWAGE PUMPS
VENTURI METER
TO PRIMARY CLARIFICATION
DOW A-23 ADDITION POINT

FIGURE 4-2 SCHEMATIC OF PRETREATMENT OF PLANT FLOW AT BENTON HARBOR - ST. JOSEPH WASTEWATER DISPOSAL PLANT

The points of chemical addition will vary depending on the particular situation. It is important, however, that addition points be downstream of return streams, such as digester supernatant, if a high degree of phosphorus removal is required.

Optimum conditions for phosphorus removal processes are best determined in two steps: a laboratory feasibility study followed preferably by full-scale plant trials if the plant exists, or by pilot trials. The feasibility study should be a short-term laboratory evaluation of various chemical systems to obtain preliminary estimates of chemicals required to meet established phosphorus removal standards. During the plant trials, chemicals and equipment should be supplied for full-scale operation over a period of 30 days or longer, to obtain detailed design data.

Chemical removal of phosphorus from wastewaters during primary treatment is a practical and controllable process which can result in low discharges of phosphorus from existing primary treatment plants and secondary treatment plants of either the trickling filter or activated sludge type. Additional costs for chemicals and attendant mixing, dispersing, and monitoring equipment are moderate.

Data and correlations given in this chapter will provide guidance and permit making design estimates. However, it would be risky to eliminate on-site experimental studies because of variations in equipment design and operation, and in wastewater characteristics.

4.2 Phosphorus Removal Data for Alum Addition

Alum has been used as a source of aluminum for phosphorus removal in raw wastewater. Sodium aluminate can also be used if the natural alkalinity of the raw wastewater is too low to prevent excessive reduction of pH by alum. However, no investigations using aluminate have been reported. Tests have been conducted on laboratory, pilot, and plant scales to determine the feasibility of alum addition to raw wastewater.

Bench-scale tests of alum addition were conducted at Springfield, Ohio (9) and Two Rivers, Wisconsin (10). All bench testing was conducted in the following manner. Samples of wastewater were stirred on a six-place stirrer for approximately 2 minutes at 100 rpm, followed by 18 minutes of gentle stirring (40 to 50 rpm). Alum additions were made during the rapid mix period. When polymers were used, they were added approximately 2 minutes after the alum addition. Treated samples were allowed to settle for 30 minutes before total phosphorus analyses were made.

The respective states have required phosphorus removals of 80% at Springfield and 85% at Two Rivers. At Springfield, an average Al:P weight ratio of 1.9:1 (molar ratio of 2.2:1) was required to reduce the average influent total phosphorus of 5 mg/l by 80%. Raw wastewater at Two Rivers required an average Al:P weight ratio of 0.93:1 (molar ratio of 1.07:1) to reduce the average influent phosphorus of 20.1 mg/l by 85%. These test data indicate that as phosphorus concentration increases, the Al:P ratio necessary to obtain a specific per cent removal decreases. Individual tests at Two Rivers, for a broad range of influent phosphorus concentrations, also showed the same variation. Influent total phosphorus ranged from 7.6 to 28.5 mg/l, and the Al:P weight ratios necessary to remove 85% ranged from 2.0:1 to 0.58:1, respectively.

A study of alum addition for phosphorus removal was conducted at Lebanon, Ohio on both pilot and plant scales (11) to determine whether chemical treatment of raw wastewater could be used successfully to precede activated carbon adsorption for organic removal.

A 15 gpm pilot plant was utilized to determine the most effective dosages of alum to be studied in the full-scale tests. Data from the pilot plant indicated that liquid alum dosages of 150, 250, and 350 mg/l, as $Al_2(SO_4)_3 \cdot 14H_2O$, and anionic polymer dosages of 0.25 and 0.5 mg/l should be evaluated. Only the Parshall flume, preaeration basin, and primary clarifiers of the 1.5 mgd Lebanon plant were used in this study. Alum was fed to the raw wastewater at the Parshall flume to assure complete mixing. The preaeration basin was used for flocculation, with polymer being added toward the end of the basin. The primary basins served as settlers for the precipitated phosphorus. Figure 4-1 shows the chemical addition arrangement at Lebanon which was used for both alum and iron studies. The iron experiments will be covered later in this chapter.

Each dosage was evaluated for five days. A polymer dose of 0.5 mg/l was found to be satisfactory for improving floc formation and settling. An alum dose of 250 mg/l, Al:P weight ratio of 3.4:1 (molar ratio of 3.9:1), removed 93% of the phosphorus in this high alkalinity (400 mg/l as $CaCO_3$) wastewater. Effluent phosphorus concentration did not exceed 0.7 mg/l and averaged 0.5 mg/l. The alum dose of 150 mg/l, Al:P weight ratio of 1.5:1 (molar ratio of 1.7:1), removed only 74% of the phosphorus; while the alum dose of 350 mg/l, Al:P weight ratio of 3.9:1 (molar ratio of 4.5:1), reduced the phosphorus by 94%. The alum dose needed to meet the State removal requirement of 80% would fall between the 150 and 250 mg/l dosages studied.

4.3 Phosphorus Removal Data for Iron Addition

Operating data from several plants utilizing mineral addition before the primary clarifier for phosphorus removal are shown in Table 4-1.

With primary treatment alone, phosphorus removals varied from 60 to 91%. In secondary plants, iron addition before the primary resulted in phosphorus removals from 75 to 93% over the entire plant. Iron doses varied between 10 and 90 mg/l as Fe. The Lebanon study showed that doubling the iron dose (from 45 to 90 mg/l) with the same polymer addition raised phosphorus removal from 68 to 91%. Polymer doses in all of the iron studies ranged from 0.26 to 0.7 mg/l.

Table 4-1

PHOSPHORUS REMOVAL OPERATING DATA FOR MINERAL ADDITION BEFORE PRIMARY CLARIFIER

Location	Flow mgd	Metal	Dose mg/l or M:P (molar)	Polymer Used	Total Phosphorus Plant Influent mg/l	Total Phosphorus Primary Effluent mg/l	Phosphorus Removal in Primary percent	Reference
Lebanon, Ohio	1.01	$FeCl_3$	45 or 0.98:1	- - -	- - -	- - -	68	11, 12
	0.86	$FeCl_3$	90 or 1.9:1	Yes	- - -	- - -	91	
Grayling, Michigan	0.3	$FeCl_2$	15-25	Yes	10-20	3-6	60-80	8, 13
Lake Odessa, Michigan	- - -	$FeCl_2$	15-25	Yes	- - -	- - -	75*-93*	14, 15
Benton Harbor, Michigan	5.5** (activated sludge return to primary)	$FeCl_3$	21 or 1.6:1	Yes	7.6	0.8*	90*	16
	5.5** (no sludge return)	$FeCl_3$	21 or 1.6:1	Yes	7.6	0.8*	65 91*	
Warren, Michigan	36	$FeCl_3$	15	- - -	- - -	- - -	80*	17
Texas City, Texas	0.7	$FeCl_2$	1.3 to 1.6:1	No	- - -	- - -	80*-90*	18
	0.7	$Al_2(SO_4)_3$	1.6 to 1.8:1	No	- - -	- - -	80*	
	0.7	$Al_2(SO_4)_3$	2.5:1	Yes	- - -	- - -	>90*	
Trenton, Michigan	4.5	$FeCl_2$	20-40	- - -	11.7	1.5	87	19
Waukegan, Illinois	15	$FeCl_3$	2.5:1	Yes	8.2	0.7	91	20
	15	$FeCl_3$	1.6:1	Yes	9.6	0.9	91	
Pontiac, Michigan	10	$FeCl_3$	15	Yes	7.0	0.9	85-90	21
	10	$FeCl_3$	16 (split)	Yes	7.2	0.5	93	
	10	$FeCl_3$	10	Yes	7.5	1.6	80	

See footnotes at end of table.

Table 4-1 (Continued)

Location	Flow mgd	Metal	Dose mg/l or M:P (molar)	Polymer Used	Total Phosphorus Plant Influent mg/l	Total Phosphorus Primary Effluent mg/l	Phosphorus Removal in Primary percent	Reference
Michigan City, Indiana	9	$FeCl_3$	14.4	...	10.0	1.7	83	22
	9	$Al_2(SO_4)_3$	7.2	...	9.0	1.0	89	
Babbitt, Minnesota	0.4	$Na_2Al_2O_4$	1.1:1	No	11.0	0.8	93	23
Dogue Creek STP, Virginia	3	$FeCl_3$	1.6:1	Yes	9.0	1.0	89	24
Little Hunting Creek STP, Virginia	4	$FeCl_3$	1.9:1	Yes	10.0	1.0	90	24
Essexville, Michigan	0.5	$FeCl_3$	17 or 2:1	Yes	4.0	0.5*	88*	25
Mentor, Ohio	4	$FeCl_2$	3.1:1	...	...	...	80	26,27

NOTE: *Values are for entire plant.
**Design Capacities: 8 mgd (primary), 4 mgd (secondary).

Iron carry-over occurred at Mentor, where $FeCl_2$ addition alone resulted in an effluent Fe concentration of 42.5 mg/l. Lime addition reduced the iron leakage to an effluent Fe concentration of 11.6 mg/l. Polymers were ineffective in reducing iron leakage.

Increased suspended solids and BOD removals were also obtained with iron additions. At Grayling, Michigan, suspended solids removal was increased by 27% to an average of 78%. At Benton Harbor, Michigan, where waste activated sludge is returned to the primary, iron addition gave a 3.0% increase in solids removal over the entire treatment plant. An example of increased BOD removal is seen in the Grayling, Michigan study. Removal of BOD in the primary was increased by 45% after iron addition. This resulted in a total BOD removal of 58% in the primary.

Typical values for primary settler detention times were 2.3 hours at Mentor and 2.25 hours at Benton Harbor. At Mentor, air at a rate of 42.5 to 85 ft^3/minute was used for flocculation in the mixing area of the primary tank.

Although lab test procedures should be adjusted to best approximate individual plant conditions, outlines of procedures used at two locations may be useful as guidelines. Jar tests were conducted at Mentor in the following manner. Iron, lime, and polymer were added one minute apart to 1,000 ml of raw wastewater at a stirrer speed of 100 rpm. The chemicals and raw wastewater were then mixed for five minutes at 20 rpm, and finally the mixture was allowed to settle quiescently for 1/2 hour before a portion of the supernatant was pipetted off for analysis.

4.4 Overall Performance of Treatment Plants with Mineral Addition to the Primary Settler

4.4.1 Primary Treatment Alone

The relative efficiencies of primary and secondary treatment with and without primary phosphorus removal are summarized in Table 4-2. Phosphorus removal without chemical precipitation in the primary clarifiers ranges from only 10 to 20% through a conventional secondary treatment plant. At least one half or more of this removal is due to sedimentation of the insoluble phosphorus fraction in the raw wastewater during primary clarification. The addition of the precipitation-flocculation process in primary clarification will result in 70 to 90% removal of total phosphorus in the primary. These removal levels can be expected only in well designed primary clarifiers that are not hydraulically overloaded. Suspended solids removals in the primaries will be increased from 40 to 70% to 60 to 75%. This increase occurs despite the additional solids generated by phosphorus removal. BOD removals in the primaries will be increased from 25 to 40% to 40 to 50%. The additional removal is primarily due to the suspended BOD fraction.

4.4.2 Effect on Secondary Treatment

Additional phosphorus removal through either trickling filter or activated sludge units following chemical precipitation in the primaries can amount to 10 to 15%. This

Table 4-2

EFFECTIVENESS OF PRIMARY AND SECONDARY TREATMENT WITH AND WITHOUT MINERAL ADDITION FOR PHOSPHORUS REMOVAL

	Phosphorus Removal (%)		Suspended Solids Removal (%)		BOD Removal (%)	
	Without	With	Without	With	Without	With
Primary Treatment	5-10	70-90	40-70	60-75	25-40	40-50
Secondary Treatment						
Trickling Filter	10-20	80-95	70-92	85-95	80-90	80-95
Activated Sludge	10-20	80-95	85-95	85-95	85-95	85-95

removal is attributed to biological flocculation. Split addition of metal in both the primary and secondary portions of the plant should be considered if exceptionally high levels (>90%) of total phosphorus removal are desired. Addition points for the secondary metal feed at or near the effluent end of the biological treatment unit should be included in the design.

Increased capture of suspended solids in the primaries results in reduced loading to the secondary units. Although the solids entering the secondary are reduced, the liquid volumes remain the same and, therefore, normal detention times should be used. Essentially the same amount of soluble BOD must be insolubilized and removed. The additional removal of suspended solids and BOD across the primaries will reduce the amount of waste activated sludge to be handled. Savings in plant operating costs may result from reduced air requirements for aeration, decreased chlorine demand, and improved sludge filterability (28).

4.5 Dosage Selection and Correlation

The material presented in Chapters 6 and 7 on choice of chemicals and dosage in trickling filter plants and activated sludge plants generally applies also to mineral addition in primary plants. Therefore, the reader is referred to these chapters for discussions of requirements for influent phosphorus data, procedures for obtaining dosage design data, calculations of daily doses, and scheduling of dose rate changes to match phosphorus variation. The reader is also referred to Chapter 10 for general information on various chemicals. Material presented here will be limited to comments and/or data on dosage selection, dosage correlation, and influent phosphorus variations.

It is generally desirable to plot or correlate test results on a simple basis to make them more useful. A reasonably good correlation can be obtained by plotting the log of the fraction of soluble phosphorus remaining as a function of the weight ratio of metal to soluble phosphorus. Typical of such plots is Figure 4-3, where both the fraction of soluble phosphorus remaining and the percent removed are plotted on the log scale. The data in Figure 4-3 are from several ferric chloride phosphorus removal studies conducted in the Great Lakes region.

Since there is usually considerable scatter in the data, such correlations should be used very conservatively as a guide for predicting doses. On-site tests under actual plant conditions are always preferred as the basis for such a plot. The required ratio of metal ion to influent soluble phosphorus for a specified phosphorus removal can be obtained from such a plot. This ratio is multiplied by the soluble influent phosphorus concentration in mg/l to determine the required chemical dose in mg/l.

As will be discussed in Chapter 6, the dose rate should be varied during the day to match more closely the varying influent phosphorus concentration. The need for doing this is illustrated by Figure 4-4 where some typical data on the diurnal variation of ortho phosphorus in plant influent are presented.

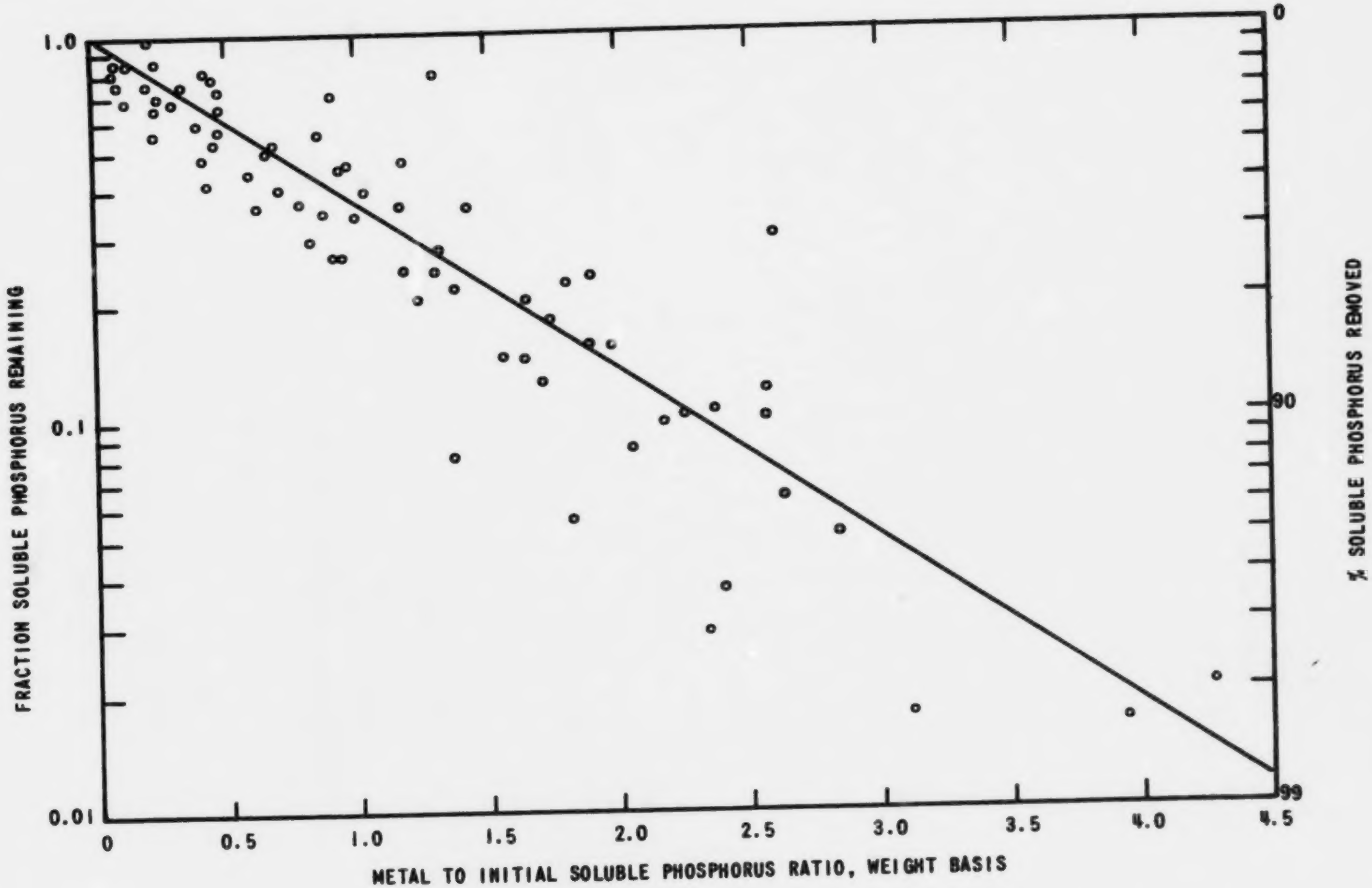


FIGURE 4-3 SOLUBLE PHOSPHORUS REMOVAL BY FERRIC CHLORIDE ADDITION

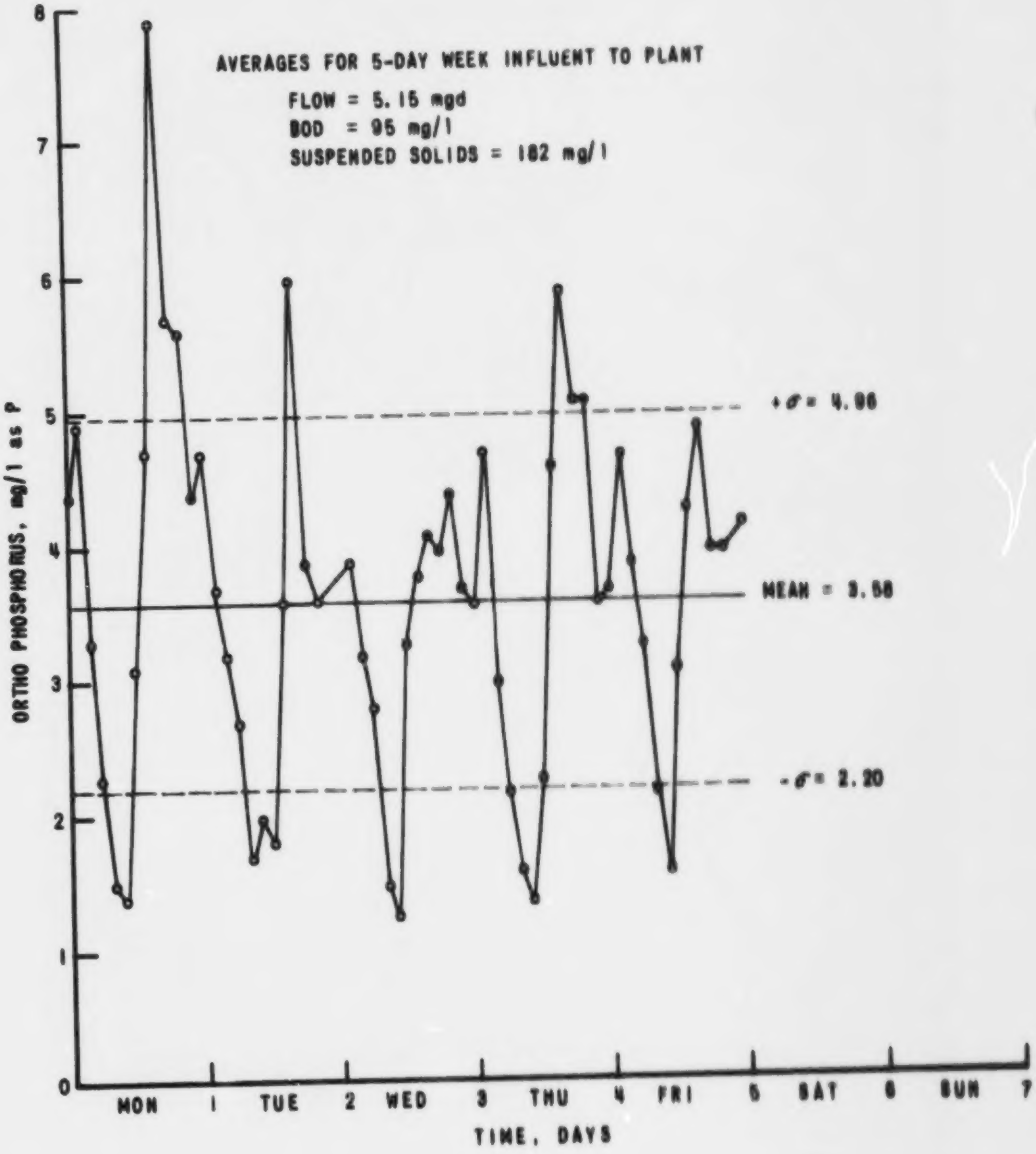


FIGURE 4-4 DIURNAL VARIATION OF ORTHO PHOSPHORUS

4.6 Case Histories

This section presents information on design and operation of three treatment plants which remove phosphorus by adding metals before primary settling and by adding polymers as required to supplement these metals.

The plant at Pontiac, Michigan employs commercial grade ferric chloride as a metal source; Trenton, Michigan uses waste pickle liquor; Waukegan, Illinois has successfully used three liquid forms of metals: ferric chloride, alum, and waste pickle liquor. Operations have been effective at all three plants, and details are presented here to illustrate the variety of approaches to phosphorus removal which are presently being applied successfully.

4.6.1 Pontiac, Michigan

The Auburn Plant in Pontiac, Michigan is a conventional activated sludge treatment plant with an average design flow of 10 mgd. The facility was constructed in 1963 according to design parameters presented in Table 4-3. Very simple chemical feed facilities were added to permit addition of ferric chloride and polymers to the primary section of the plant as shown in Figure 4-5.

A test program in 1969 (21) showed that feeding ferric chloride and polymer ahead of the primary clarifier was an effective procedure to remove 80 to 90 percent of total phosphorus in the wastewater. Long-term analysis of concentrations and forms of phosphorus in plant influent indicate soluble compounds represent 60 to 80 percent of the total phosphorus present, and that orthophosphates represent 35 to 45 percent of total phosphate. There appeared to be minimal iron leakage in the effluent, and overall plant performance, in terms of BOD and SS removals was improved by this technique.

The best point for addition of ferric chloride was just before the aerated grit chamber at the head of the plant. Theoretical displacement through the grit chamber was approximately 10 minutes, but observations indicated actual detention was more on the order of 2 to 5 minutes at average flow rates. This was sufficient time, however, for mixing and initial flocculation of the metal salt. Passage through the grit unit provided adequate lag time before polymer addition.

Since 1972, the plant has been operating approximately as shown by the performance data in Table 4-4 (29).

These data, mostly derived from 1973 experience, reflect a high-quality effluent and well-stabilized sludge. When feeding chemicals, the primary section of the plant removes 40 to 60 percent of incoming BOD, 50 to 75 percent of incoming SS, and 50 to 75 percent of incoming total phosphorus.

Table 4-3

DESIGN DATA FOR PONTIAC, MICHIGAN

Description	
Design Flow, mgd	
Average Dry Weather	10
Peak Dry Weather	20
Maximum	36
Primary Clarifiers	
Detention Time,[1] hr	1.5
Overflow Rate,[1] gpd/sq ft	1,000
Sludge Solids Concentration, percent	6.5
Sludge Volume, percent of total flow	0.08
Aeration Tanks	
Detention Time,[1] hr	6.5
MLSS, mg/l	2,000-3,000
Return MLSS, percent	40
Final Clarifiers	
Detention Time,[1] hr	1.9
Overflow Rate,[1] gpd/sq ft	790
Sludge Digesters	
Total Active Volume, cu ft	195,000
Vacuum Filters	
Total Area, sq ft	375
Nominal Capacity, psf/hr	4

[1] Based on Average Flow

There has been no evidence of serious problems related to escape of colloidal iron from the treatment system. Some 2 to 3 mg/l of soluble or colloidal iron is found both in plant influent and effluent.

Both ferric chloride and polymer are needed in this operation for good overall performance. If iron dosage is reduced below the indicated mole ratio of 1.7, or if polymer addition is discontinued, effluent quality begins to degrade in a matter of a few hours.

FIGURE 4-5 TREATMENT FACILITIES AT PONTIAC, MICHIGAN

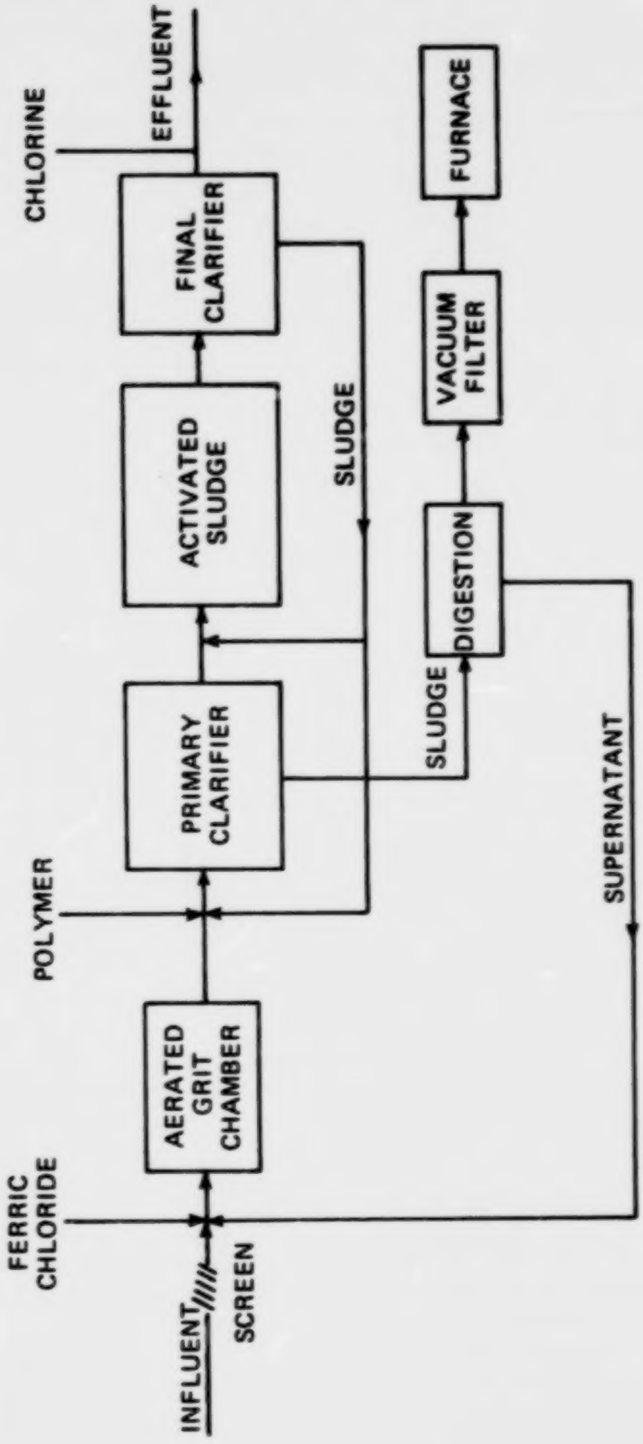

INFLUENT
SCREEN
FERRIC CHLORIDE
AERATED GRIT CHAMBER
POLYMER
PRIMARY CLARIFIER
SLUDGE
DIGESTION
SUPERNATANT
ACTIVATED SLUDGE
SLUDGE
FINAL CLARIFIER
CHLORINE
EFFLUENT
VACUUM FILTER
FURNACE

Table 4-4

OPERATING DATA FOR PONTIAC, MICHIGAN

Description	
Flow, mgd	10
Influent Characteristics	
BOD, mg/l	100
SS, mg/l	140
Total P, mg/l	5
Chemical Addition	
Iron Dose, mg/l	15
Iron Dose, Fe/P	1.7
Polymer Dose, mg/l	0.3
Overall Plant Performance	
Effluent BOD, mg/l	10
Effluent SS, mg/l	20
Effluent Total P, mg/l	1
BOD Removal, percent	90
SS Removal, percent	86
Total P Removal, percent	80
Digesters	
Temperature, deg F	90
Solids Loading, lb VSS/cu ft/day	0.09
Digested Sludge Concentration, percent	7.2
Digested Sludge Volatile Fraction, percent	45
Vacuum Filters	
Loading, psf/hr	3-4
Cake Solids Concentration, percent	30-35

Volume of sludge produced by chemical addition is approximately the same as generated during conventional treatment. The chemical sludge is more concentrated but there have been no problems with digestion or dewatering of sludges produced in this location. Digested sludge is conditioned with ferric chloride and lime (at about the same dosage rates as previously required for conventional sludge treatment) prior to vacuum filtration. Cake forms readily on the filters but at modest yield rates. Both the vacuum filters and the multiple hearth furnace which follows are operated continuously for periods of 60 to 75 hr during midweek. The units are shut down, cleaned, and held on standby the rest of the time.

Cost of treatment chemicals in 1973 was 2¢/1,000 gal of wastewater treated. Of this cost, 85 percent was for ferric chloride. Exact cost records are not available on capital equipment installed in this system. That modest equipment outlay consisted of fiberglass tanks and chemical metering pumps. Although quite workable, this temporary chemical feed arrangement is to be replaced by permanent improvements in a 1974 plant expansion.

4.6.2 Trenton, Michigan

Since December 1970, Trenton, Michigan has operated a 5.5 mgd activated sludge treatment plant with addition of pickle liquor and polymer for phosphorus removal and effluent polishing (19,30). Design data for the Trenton plant are listed in Table 4-5.

Pickle liquor, the only metal salt used on a long-term basis at Trenton, is stored in two rubber-lined steel tanks having a capacity of 25,000 gal each. These tanks have hot water heating coils to keep coagulant temperature high enough to prevent crystallization of solids during cold weather. The storage tanks are filled by tank trucks bringing pickle liquor from a nearby steel plant; the trucks discharge into the storage tanks by applying low pressure compressed air. The liquid metal salts are fed, by gravity, in proportion to the rate of wastewater flow through the plant at a preselected concentration of metal. The concentration can and is changed from time to time to meet changing needs.

Two polymer tanks, each holding 500 gal, are outfitted with variable flow metering pumps which can be manually set to predetermined rates. These inject polymer between the activated sludge aeration tankage and the final clarifiers. Mechanical mixing equipment is provided in a small box between these two operations, and medium energy mixing is provided there to blend polymers into the main flow.

Predesign pilot scale testing and plant scale startup operations indicated the system performed best in the mode illustrated by Figure 4-6. Pickle liquor was added just ahead of the aerated grit chamber, and the polymer was added to the activated sludge effluent en route to the final clarifier.

Both the polymer addition and tube settlers installed in the final clarifier were intended to improve collection and capture of fine floc observed in primary effluent during studies in the late 1960's.

Table 4-6 shows performance data typically observed at this treatment facility since it has been in full scale service.

Subsequent to these operating data, an additional inflow of industrial waste amounting to 1 mgd has been added to the plant flow. The treatment system is presently accepting this

Table 4-5

DESIGN DATA FOR TRENTON, MICHIGAN

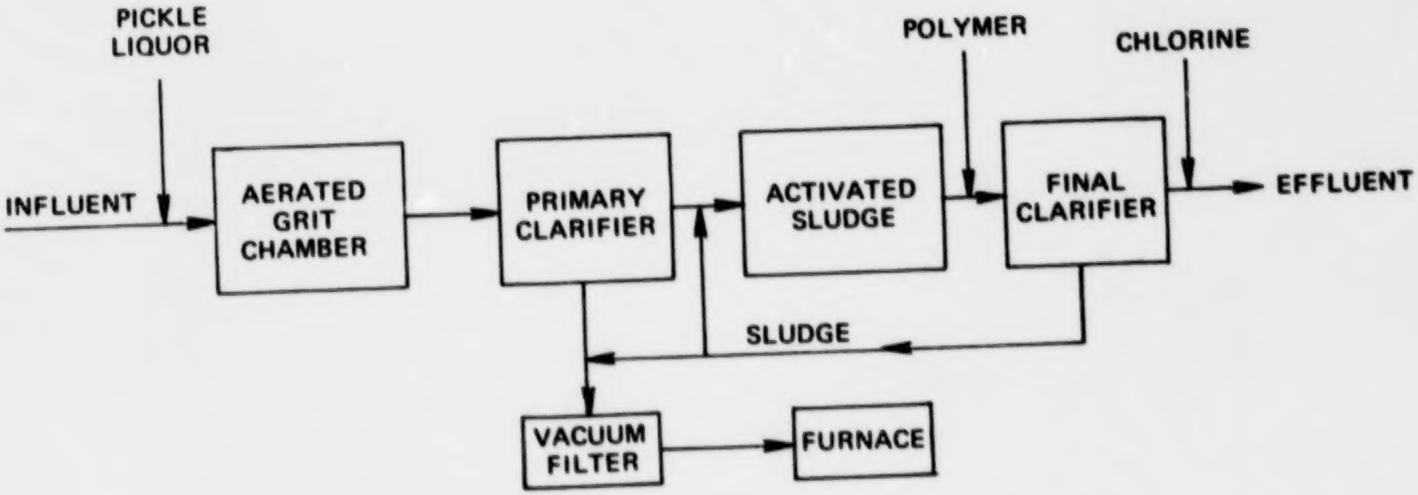
Description	
Design Flow, mgd	
Average Dry Weather	5.5
Peak Dry Weather	6.5
Maximum Rate	16.5
Primary Clarifiers	
Detention Time,[1] hr	1.3
Overflow Rate,[1] gpd/sq ft	1,440
Weir Rate,[1] gpd/ft	23,000
Aeration Tanks	
Detention Time,[2] hr	5.5
Return Flow, percent of average	30
Final Clarifiers (Tube Module Type)	
Detention Time,[1] hr	3.5
Tube Overflow Rate, gpd/sq ft	1,200
Total Surface Overflow Rate, gpd/sq ft	500
Sludge Production	
Primary (dry), lb/day	4,000
Secondary (dry), lb/day	5,000
Chemical (dry), lb/day	4,000
Primary plus Chemical, percent	5
Secondary, percent	1.0-1.5
Gravity Thickening	
Solids Loading, psf/day	6.2
Vacuum Filtration	
Area, sq ft	360
Nominal Capacity, psf/hr	2-3

[1] Based on Average Flow

[2] Including Return Flow

FIGURE 4-6

TREATMENT FACILITIES AT TRENTON, MICHIGAN


PICKLE LIQUOR
INFLUENT
AERATED GRIT CHAMBER
PRIMARY CLARIFIER
ACTIVATED SLUDGE
POLYMER
FINAL CLARIFIER
CHLORINE
EFFLUENT
SLUDGE
VACUUM FILTER
FURNACE

additional waste, plus other increased domestic loads, so that average daily flow is now 7.5 mgd. This additional waste includes an increase in BOD loading to the plant of about 4,000 lb/day. Plant performance at this increased load is reported as essentially the same as in Table 4-6. Preliminary operating data from early 1974 indicate removal of 90 percent of the BOD and SS, and 80 percent of total phosphorus. Sludge processing has continued on a stable basis as indicated in Table 4-6.

Total cost of the 1974 facilities was $4.8 million. In this major construction project, it was not possible to separate the relatively minor expenses related to the chemical treatment facilities themselves. Something on the order of $5,000 could be identified for expenditures

Table 4-6

OPERATING DATA FOR TRENTON, MICHIGAN

Description	
Flow, mgd	5.5
Influent Characteristics	
BOD, mg/l	220
SS, mg/l	300
Total P, mg/l	9.5
Ortho P, mg/l	5.1
Chemical Addition	
Iron Dose, mg/l	20
Iron Dose, Fe/P	1.2
Polymer Dose, mg/l	0.4
Overall Plant Performance	
Effluent BOD, mg/l	10
Effluent SS, mg/l	10
Effluent Total P, mg/l	1
Effluent Ortho P, mg/l	0.5
BOD Removal, percent	95
SS Removal, percent	97
Total P Removal, percent	89
Ortho P Removal, percent	90
Gravity Thickeners	
Thickened Sludge Concentration, percent	5.1
Thickened Sludge Mass, lb/day	16,500
Cake Solids Concentration, percent	15-20

for specific hardware, but concrete, electrical service, and other such features could not be realistically identified as if they were an independent capital improvement. At any rate, facility costs would be very low when capitalized over a period of years.

In 1973, the total operating cost of this facility was approximately 20¢ to 22¢/1,000 gal of wastewater treated; this cost included all phases of wastewater plant operation. Operating cost for chemical addition was quite low, on the order of 1¢/1,000 gal, consisting of about 0.5¢/1,000 gal for polymer, and slightly less than that for the very economical source of iron represented by pickle liquor. Iron in pickle liquor analyzed 0.75 lb of iron/gal and the delivered cost was 2¢/gal. Therefore, delivered cost of iron was 2.7¢/lb.

4.7 Costs

All cost estimates for the removal of phosphorus by chemical precipitation-flocculation during primary treatment are dependent to some extent on raw wastewater flow rates. The cost per 1,000 gal. for chemicals may be less in larger installations because of large volume purchasing at lower unit prices. Total capital costs for tankage, chemical feed pumps, and instrumentation will be less for smaller plants. Placed on a unit volume of waste treated basis, however, capital costs for small plants will be higher than for large plants.

4.7.1 Chemical Costs

The chemical costs for phosphorus removal in the primary settler have been computed for both liquid alum and liquid $FeCl_3$. The respective unit costs of these chemicals were estimated at $60/ton for alum (on a 17.1% Al_2O_3 basis) and $90/ton, as $FeCl_3$, for liquid $FeCl_3$. These figures represent average delivered prices.

As a basis for estimating costs in terms of ¢/1,000 gal., an influent P concentration of 10 mg/l was assumed. For removal of 80% of the total phosphorus in primary treatment, equal molar ratios of Al:P and Fe:P of 1.5:1 were assumed. Polymer costs were included, based on a dosage rate of 0.5 mg/l and a chemical cost of $1.50/lb. The chemical costs of phosphorus removal treatment are estimated to be 3.6 ¢/1,000 gal. for liquid $FeCl_3$ and 4.2 ¢/1,000 gal. for liquid alum, respectively. As mentioned earlier, these costs may vary depending on the size of treatment plant and cost of freight. Actual costs should be determined prior to detail design.

The chemical cost for phosphorus removal at plants where waste pickle liquor is available is more difficult to predict accurately because of varying iron concentrations, larger storage requirements, freight charges, and the additional alkalinity requirement.

4.7.2 Capital Costs

Equipment costs for phosphorus removal with aluminum and iron are small relative to the chemical costs. Costs for facilities will be presented in Chapter 10 for single point addition of several different chemicals.

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Chapter 5

PHOSPHORUS REMOVAL BY LIME ADDITION BEFORE THE PRIMARY SETTLER

5.1 Introduction

Use of lime to treat raw wastewater dates back to the mid-1800's. Observers recognized that calcium formed a precipitate both with hardness and with phosphates. Under one patented technique both phosphates and lime (CaO) were added at the treatment plant to increase the amount of precipitate formed and to aid removal of SS (1).

More recent experimental efforts involved lime treatment of raw wastewater to remove phosphorus. These are typified by an investigation in New York, in which feeding 300 mg/l of slaked lime to a typical domestic wastewater removed 94 percent of the phosphorus to an effluent concentration of 0.6 mg/l, reduced SS by 95 percent to 4 mg/l, and reduced BOD by 71 percent to an average of 101 mg/l (2).

Other studies have suggested that lime addition to primary treatment is an efficient technique (3, 4). These studies were at bench-, pilot- and plant-scale and total phosphorus removals obtained were 80 percent or more with concurrent reduction in BOD of 60 to 70 percent. In this work, sufficient lime was added to bring primary effluent to a pH of 9.5 or 10. This pH level was tolerated without problems in the activated sludge process which followed; carbon dioxide generated by biological activity served to correct pH to near neutral levels. Recirculation of lime sludge was very beneficial. With sludge reseeding, as much as 60 to 75 percent reduction in overall lime requirements were obtained compared to tertiary lime treatment. Reaction time required to achieve settleable lime-phosphorus compounds was reduced from 1.0 hr without recirculation, to 0.25 hr with reseeding.

A number of other studies, from bench tests through plant-scale trials, helped to establish lime addition to primary treatment as a viable alternate for phosphorus removal operations (5, 6, 7).

The low-lime system described in Section 5.2.1 can provide 80 percent phosphorus removal on a consistent basis. By adding tertiary filtration and facilities for addition of metal coagulants and polymer, the technique can reduce total effluent phosphorus to near 1.0 mg/l.

The high-lime system described in Section 5.2.2 offers an even higher level of efficiency. With flocculant aids and filtration, it can consistently produce final phosphorus levels below 1.0 mg/l.

5.2 System Layout and Operation

Two basic systems are considered in this section. One involves relatively low lime dosages to primary influent, with biological treatment serving both to recarbonate the wastewater and remove additional phosphorus for metabolic needs. The other approach entails high-lime treatment in the primary section, followed by pH adjustment with supplemental carbon dioxide and biological treatment. The two systems are discussed separately in this section, after the comments below which apply to both approaches.

Sludge disposal during lime addition will be more complex than in a conventional biological plant. Lime sludge may not be compatible with existing disposal processes. Total sludge weight may be doubled or tripled over conventional operation. Recovery of lime by recalcining may be less attractive in this process because treatment in primary units yields a sludge which is high in concentration of organic and extraneous inorganic materials, and relatively low in recoverable lime compounds.

Lime treatment appeals most where incoming hardness and alkalinity are low because less lime is needed and less sludge is produced. However, turbid effluent may develop and settling aids may be required to supplement bioflocculation to insure a clear final effluent (8). Recirculation of lime sludge has proven beneficial to the basic reaction taking place. It promotes a more complete reaction, reduces scaling, and lowers lime demand.

Removal of phosphorus by lime addition in primary treatment is accompanied by increased BOD and SS removal at that point. This reduces organic load on secondary facilities and could aid in establishing nitrification.

In evaluating lime treatment systems, the effect of reduced carbon loads on their activated sludge unit should be considered. Since lower organic loadings to the secondary treatment units result in less cellular synthesis, a reduction in sludge wasting may be required. However, if the sludge is retained in the aeration system (in an attempt to keep MLSS up to some prescribed level) sludge age may extend to the point that settleability suffers and phosphorus escapes by cell lysis.

Also, lime precipitation in the primary clarifier will increase the capture of silts or other similar inorganic solids. These weighting agents will no longer be carried into the activated sludge unit for sorption onto mixed liquor solids, and this may result in a fluffier biological floc.

5.2.1 Low-Lime Treatment

The basic appeal of low lime addition in primary treatment stems from two fundamentals: the chemical law of mass action and the need for phosphorus in biological systems.

In the first of these, lime precipitates more phosphorus during initial stages of their mutual reaction than it does when pH has been elevated and phosphorus concentrations are quite low. This effective performance by lime usually continues through the first 75 percent of phosphorus reduction, and that occurs before pH 10 when lime demands for softening reactions increase. Overall lime demands become much higher when the process is operated to reduce phosphorus to very low levels.

Following initial phosphorus removal by lime, in the primary facilities, biological treatment removes more phosphorus as bacteria utilize it to support cell synthesis. Generally, phosphorus is not a limiting nutrient in domestic biological treatment systems as long as primary effluent contains at least 0.2 mg/l of soluble phosphorus. Microorganisms demand phosphorus to synthesize cells in direct response to the carbon concentration of the incoming waste; in municipal wastewater, carbon is the limiting element, not phosphorus. The result is lime precipitation can proceed at economical levels, then biological treatment can further reduce phosphorus residuals to an acceptable effluent concentration.

The basic low-lime treatment system is shown schematically in Figure 5-1. One advantage of this process is the opportunity to use existing primary clarifiers for phosphorus removal. Capital expenditure, therefore, could be relatively small. The process requires equipment for feeding and flash mixing lime. Flocculation usually occurs in the inlet zone of the clarifier but separate facilities may be preferred for this process. The elevated pH of primary effluent is reduced by recarbonation due to carbon dioxide produced in biological metabolism. The low-lime system may not be appropriate at trickling filter plants unless high effluent pH is tolerable, since recarbonation would be minimal in trickling filter units.

5.2.2 High-Lime Treatment

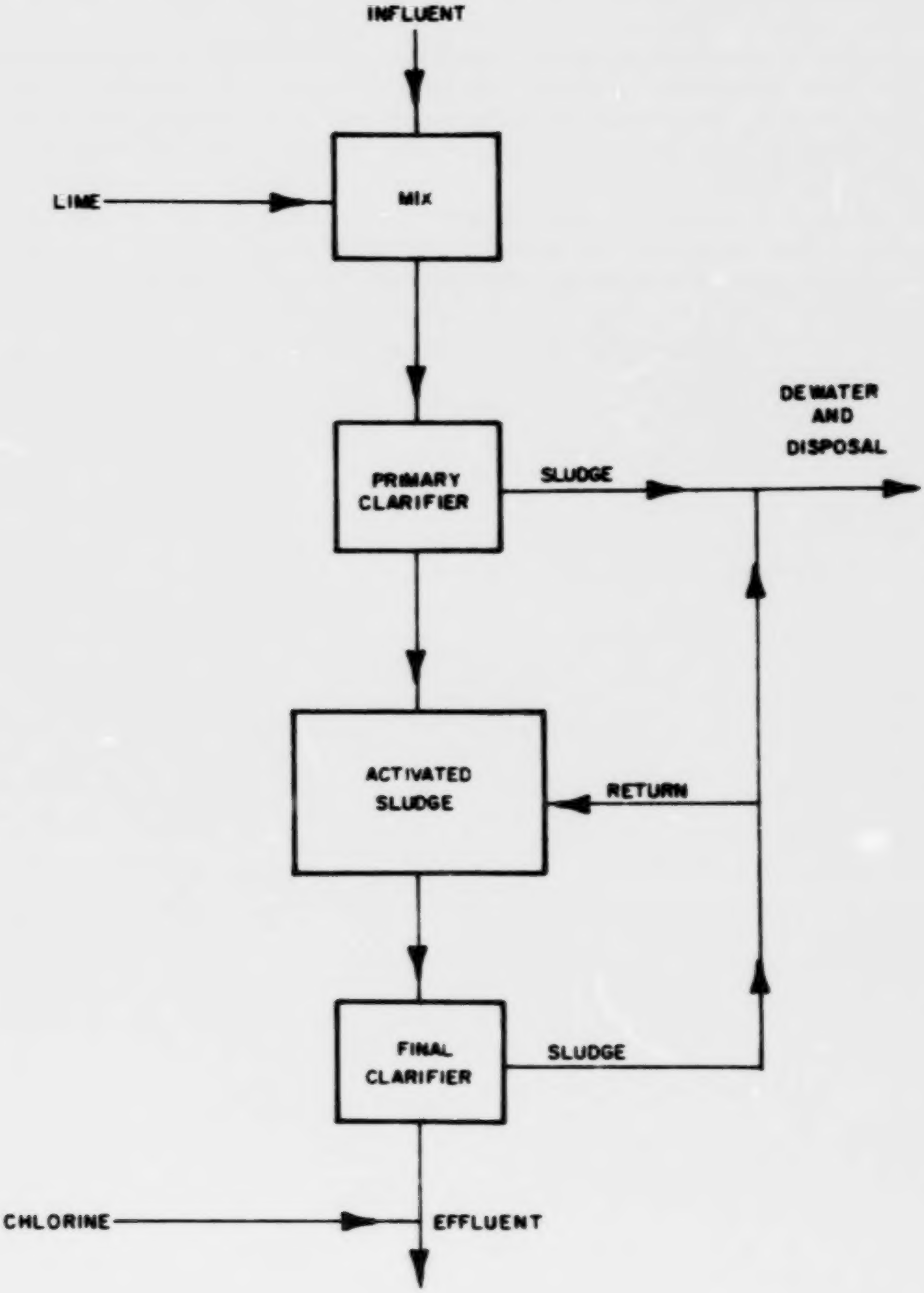
For the purposes of this manual, high-lime treatment is defined as the addition of sufficient lime in primary facilities to achieve pH 11. The basic system layout is generally as shown in Figure 5-1, for low-lime treatment. Recarbonation with stored carbon dioxide may be necessary and facilities for this would normally be provided. Recovery of lime by recalcination, an alternate feature discussed later in this Chapter, would frequently be included in high-lime systems.

High-lime treatment is applicable when effluent quality requirements include special provisions such as softening (for reuse), low levels of soluble compounds of metals, improved virus removal, or reliable and consistent reduction of phosphorus below 1.0 mg/l. Some form of tertiary solids removal, such as filtration, would usually be employed in these plants.

High-lime treatment would normally be combined with other biological, physical or chemical processes to provide an overall system of advanced waste treatment. If biological treatment is included, the primary clarifier phosphorus residual should be high enough to meet metabolic requirements.

High-lime treatment may result in a net increase in plant effluent levels of total dissolved solids and alkalinity. See Section 5.3.2.

FIGURE 5-1
TYPICAL FACILITIES FOR LOW-LIME ADDITION IN PRIMARY TREATMENT


INFLUENT
LIME
MIX
PRIMARY CLARIFIER
SLUDGE
DEWATER AND DISPOSAL
ACTIVATED SLUDGE
RETURN
FINAL CLARIFIER
SLUDGE
CHLORINE
EFFLUENT

5.3 Case Histories

Two case histories are included in this section. One involves low-lime addition in primary treatment, while the other is a high-lime system. Both installations are further cited in Section 5.5, and it is intended that information in this section will give better perspective to those later comments on sludges involved.

5.3.1 Newmarket, Ontario–Low-Lime Process

Lime fed into the primary section of an existing 2.4 mgd conventional activated sludge treatment plant in Newmarket was the first permanent installation for phosphorus removal in Ontario, Canada (5, 9). Phosphorus removal facilities have been operational in Newmarket since March, 1971.

The original treatment facility (constructed in 1963) included screening and grit removal in an aerated tank with 5 to 10 minutes detention. The original design data and subsequent operating data are shown in Table 5-1. A schematic of the Newmarket treatment facility is shown in Figure 5-2.

Table 5-1

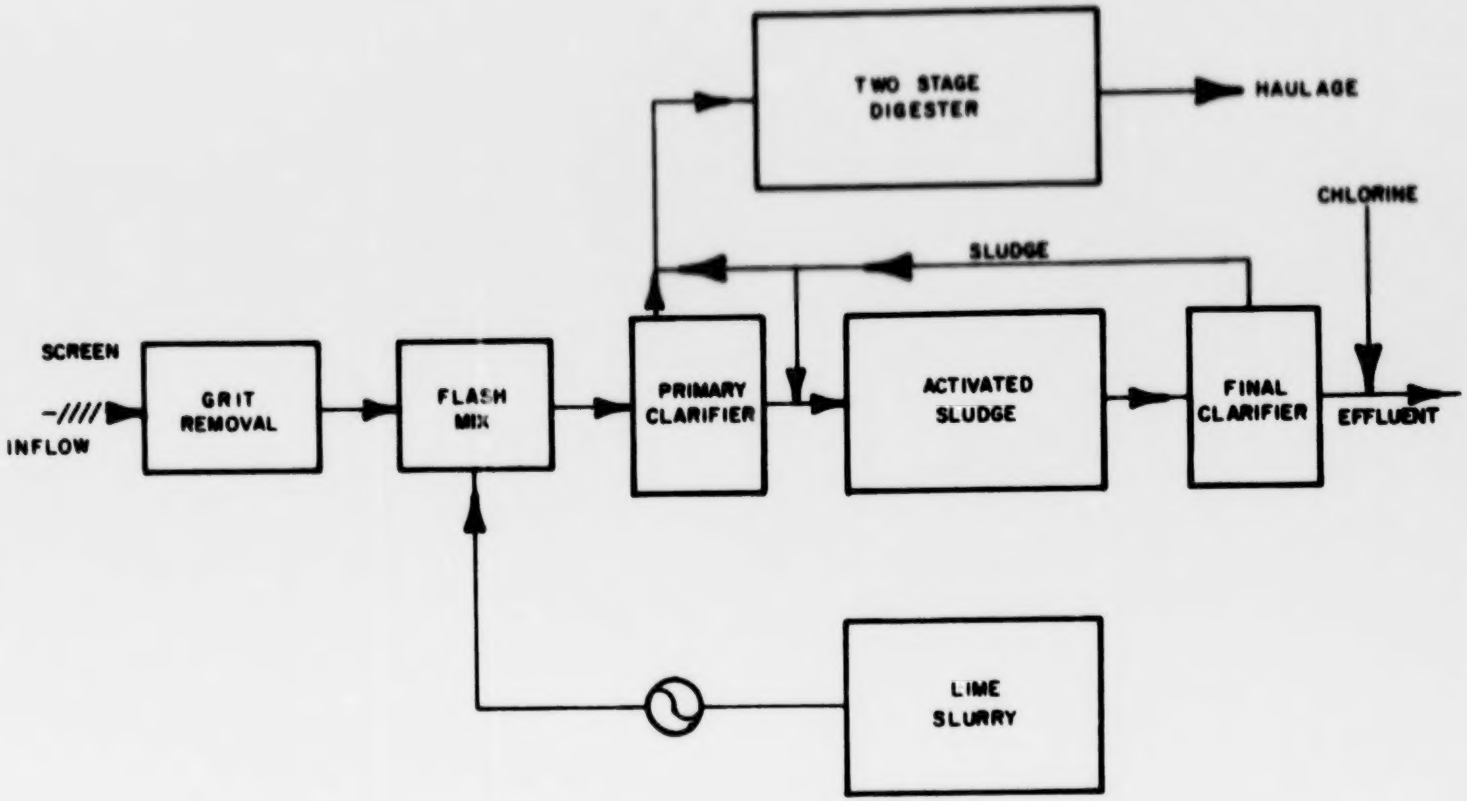
DESIGN AND OPERATING DATA FOR NEWMARKET, ONTARIO

Operation	Design Data	Operating Data
Flow, mgd	2.4	1.9
Primary Clarifiers		
Detention Time, hr	1.6	2.0
Overflow Rate, gpd/sq ft	1,330	1,050
Aeration Basins		
Detention Time, hr	8	10
F/M, lb BOD/day/lb MLSS	1.0	0.8[1]
Final Clarifiers		
Detention Time, hr	2.4	3.0
Overflow Rate, gpd/sq ft	1,000	800
Chlorine Contact		
Detention Time, hr	0.42	0.53

[1] When lime addition was established, efficient primary clarification reduced F/M to 0.2 lb BOD/day/lb MLSS; concurrently, one aeration basin was shut down so effective aeration time was reduced to 6.7 hr.

FIGURE 5-2

TREATMENT PLANT AT NEWMARKET, ONTARIO


SCREEN
INFLOW
GRIT REMOVAL
FLASH MIX
PRIMARY CLARIFIER
TWO STAGE DIGESTER
HAULAGE
SLUDGE
ACTIVATED SLUDGE
FINAL CLARIFIER
CHLORINE
EFFLUENT
LIME SLURRY

Sludge treatment includes anaerobic digestion of combined primary and waste activated sludges. The heated primary digester (40 ft. in diameter and 21.3 ft. deep) has a capacity of 2,000,000 gal and gas mixing is employed to distribute the daily solids loading of 0.1 lb/cu ft. A secondary digester, of similar size, completes the digestion process. Ultimate disposal is by wet spreading on farmland.

Lime storage is in a 25 ft diameter by 17 ft deep tank (mostly underground) which holds 30 tons of hydrated lime made up to near a 30 percent slurry by weight. To put lime into storage, bulk calcium hydroxide is discharged directly to the storage tank along with slurry water. The new slurry, after thorough blending, is sampled and titrated to determine actual lime concentration. This type of storage also allows use of waste carbide slurry from a local acetylene plant. Continuous slurry agitation is provided by either of two centrifugal slurry pumps mounted beside two positive displacement lime feed pumps. Slurry pumps require double mechanical seals for this severe service. Lime feed lines require flushing connections.

Lime feed is based on preselected dose rates with automatic interlock to rate of wastewater inflow. Metered slurry is discharged into a new flash mix box, where inflow is mechanically agitated for about 8 minutes.

Lime operating data reported in Table 5-2 is based on a 200 mg/l slaked lime dosage. Overall phosphorus removal at this dosage was greater than 80 percent.

Performance was slightly less effective at a slaked lime dose of 175 mg/l. Also, waste carbide lime, at rates equivalent to 150 mg/l of slaked lime, were reasonably effective in removing phosphorus. However, the acetylene waste proved highly abrasive to pumps and also formed heavy encrustations enroute to the biological unit. Carbonate scale problems were fewer using slaked lime, but occasional hosing down of primary launders was still necessary. Encrustation did not occur in aeration basins.

Discharge of relatively alkaline (average pH 9.6 with peaks to pH 11.5) primary effluent to the activated sludge system does not appear to distress the biological process.

Waste activated sludge is returned to the primary clarifiers and the resulting sludge mixture is drawn to anaerobic digestion. During conventional operation about 5 gal total sludge/1,000 gal plant flow was produced. Primary underflow was about 4 percent solids, of which 65 percent were volatile.

During lime addition, waste activated sludge production is reduced 50 percent due to the lower organic loads on secondary treatment. Additional sludge from lime reactions keep primary underflow at 5 gal/1000 gal plant flow. In this mode, however, the solids content averages 9 percent or more, of which 36 percent are volatile. The pH ranges from 8 to 10 in this sludge. Concentrations up to 15 percent solids were obtained when operators thickened in situ by allowing sludge depth to build to 2.5 ft on the clarifier bottom. No problems developed with mechanical scrapers, and sludge lines remained unplugged.

Table 5-2

OPERATING RESULTS FROM NEWMARKET, ONTARIO

	Conventional Treatment				Lime Treatment			
	Influent	Primary	Final	Percent Removal	Influent	Primary	Final	Percent Removal
BOD	249	172	21	92	227	84	9	96
SS	341	126	17	95	321	80	7	98
Total Phosphorus	14.2	12.5	8.7	39	10.3	2.7	2.0	81
Orthophosphate	8.1	8.3	7.6	6	6.7	1.2	1.8	73
Hardness	225	229	235	-	239	240	277	--
Alkalinity	386	377	227	-	391	374	247	-
pH	7.8	7.8	7.7	-	8.0	9.6	8.0	-

Note: All data are mg/l except pH: hardness and alkalinity are expressed as calcium carbonate.

Chemical-biological sludge is pumped to digestion, as in prior conventional operation. The anaerobic system has adapted to this type sludge and complete digestion is obtained. Operation over a period of months has indicated that digester stability was enhanced when sludge was thickened by holding a sludge blanket nearly 2.0 ft deep in the primary clarifier.

This brought sludge pH down to 9.5 or less on a consistent basis, and thickened the mass to 11 percent solids. Digested sludge has a pH near 7.2. Digestion produces normal amounts of methane gas, but free carbon dioxide is reduced by consumption in recarbonation reactions. Digester supernatant contains less than 5 mg/l of soluble phosphorus.

Capital costs, in 1971 dollars, for the lime treatment additions are shown in Table 5-3. Operating costs are almost entirely limited to the 200 mg/l lime consumption. In this instance, there is no increase in the cost of sludge handling because the volume of digested sludge has remained unchanged.

5.3.2 Central Contra Costa Sanitary District (CCCSD), California–High-Lime Process

Recent plant-scale tests of lime addition before primary treatment have been conducted in Contra Costa County, California (10,11,12). A complete advanced treatment test facility has been in operation since November 1971. Full-scale units provide tests of all elements of a 30 mgd water reclamation system currently under construction and scheduled for start-up in 1976. Since the purpose of the CCCSD water reclamation system is to reclaim wastewater for industrial reuse effluent requirements are quite strict and are listed in Table 5-4.

Table 5-3

CONSTRUCTION COSTS AT NEWMARKET, ONTARIO

Structures	$23,000
Mechanical Work	20,000
Electrical Work	8,000
Equipment	5,000
Engineering	9,000
Miscellaneous	4,000
Total	$69,000

Table 5-4

EFFLUENT REQUIREMENTS FOR CCCSD TREATMENT FACILITY

Constituent	Limit (mg/l)
Total Phosphorus	1.0
BOD	10.0
Total Nitrogen	5.0
Hardness (as $CaCO_3$)	300
Alkalinity (as $CaCO_3$)	255
TDS (maximum increment above water supply)	375

In order to demonstrate effectiveness of the planned 30 mgd treatment system, investigators converted existing treatment facilities to various plant-scale forms of advanced waste treatment. The final layout of the overall test facility is shown in Figure 5-3. Table 5-5 gives pertinent data.

After screening, raw sewage passed a chemical addition point where lime was added and flash mixed with recycled primary sludge. Ferric chloride and polymer addition were also possible at the same point. Wastewater then flowed to the preaeration basin. Heavy grit

FIGURE 5-3

SYSTEM LAYOUT AT CONTRA COSTA COUNTY

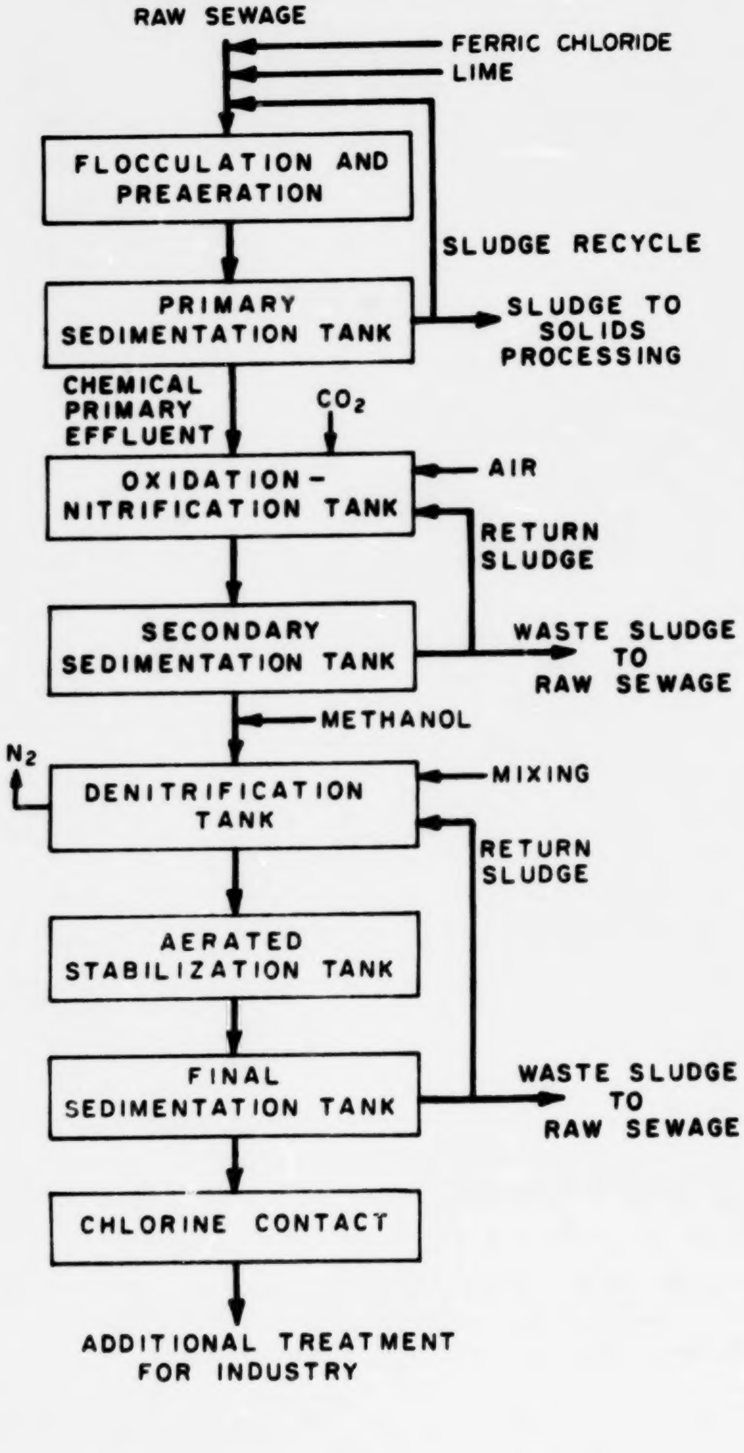

RAW SEWAGE
FERRIC CHLORIDE
LIME
FLOCCULATION AND PREAERATION
SLUDGE RECYCLE
PRIMARY SEDIMENTATION TANK
SLUDGE TO SOLIDS PROCESSING
CHEMICAL PRIMARY EFFLUENT
CO_2
AIR
OXIDATION-NITRIFICATION TANK
RETURN SLUDGE
SECONDARY SEDIMENTATION TANK
WASTE SLUDGE TO RAW SEWAGE
METHANOL
N_2
DENITRIFICATION TANK
MIXING
RETURN SLUDGE
AERATED STABILIZATION TANK
FINAL SEDIMENTATION TANK
WASTE SLUDGE TO RAW SEWAGE
CHLORINE CONTACT
ADDITIONAL TREATMENT FOR INDUSTRY

Table 5-5

BASIC DATA FOR CCCSD

Operation	
Preaeration and Grit Removal	
Average Water Depth, ft	11
Detention Time, hr	0.53
Design Flow/Tank, mgd	2.5
Air Supplied, cu ft/gal	0.10
Primary Clarification	
Average Water Depth, ft	10
Detention Time, hr	1.6
Overflow Rate, gpd/sq ft	1,200
Design Flow/Tank, mgd	2.5

particles such as sand and unreacted lime settled out there. The gentle air agitation following lime addition promoted the growth of large floc particles which were readily settleable. Long-term operation of this air mixing facility indicated it was effective, and this approach eliminated problems associated with mechanical flocculators in raw wastewater.

Waste activated sludge was returned to the preaeration section; there it mixed and settled with the chemical and primary sludges. The system had no separate facilities for treatment of waste activated sludge.

In the preaeration basin it appeared the carbon dioxide content of agitation air might, by the process of recarbonation, work against the lime reactions required in the preaeration unit. However, if all carbon dioxide in the air were transferred into solution, less than 0.5 mg/l dissolved carbon dioxide would be introduced. This was considered insignificant.

However, carbon dioxide in the preaeration air did cause scaling inside the orifices of the coarse bubble diffusers. Initially the tank had to be drained every week or two to clean plugged orifices. Installation of high-pressure air blowoff facilities extended that period to two months. In the 30 mgd plant, swing diffusers will be provided so tanks will not have to be taken out of service for diffuser maintenance.

The lime-treated effluent was then settled in the rectangular primary tanks under operating parameters described previously in Table 5-5.

During the course of the studies lime dosage was varied, and supplemental coagulants were utilized to modify effluent qualities. Data for various modes of operation are shown in Tables 5-6 and 5-7.

Operation at pH 11 was not as effective in phosphorus removal as operation at 11.5 (see Table 5-5). The phosphorus remaining in the primary effluent at pH 11 was 2.3 mg/l compared with less than 1 mg/l at the higher dose.

Ferric chloride was tested as a supplemental coagulant to enhance capture of colloidal phosphorus precipitates. At pH 11, a ferric chloride dose of 14 mg/l produced primary effluent phosphorus of less than 1 mg/l (see Table 5-6). Even the relatively low lime dose for pH 10.2 was possible if compensated by ferric chloride dosage of 24 mg/l. In this operation primary effluent total phosphorus averaged 0.68 mg/l.

Removals of BOD, SS, and other parameters were comparable in each mode of operation. However, increase in hardness across the primary varied considerably. Operation at pH 11 with ferric chloride yielded the lowest effluent hardness.

The data in Tables 5-6 and 5-7 represent results when primary clarifier underflow solids were recycled to the chemical addition point. This was done to improve precipitation reactions involving calcium, magnesium, and phosphorus. Total solids in the lime reactor were maintained in the range of 1,000 to 2,500 mg/l. Later with solids recycle discontinued, the major performance change was in effluent hardness. During operation at pH 11 with ferric chloride dose of 10 to 16 mg/l, hardness across the primary increased an average of 32 mg/l. Under similar chemical feed rates, hardness increase across the primary was 6 mg/l when flocculator solids were in the range of 900 to 1,500 mg/l TSS. When flocculator solids were increased to 1,700 to 3,900 mg/l, hardness increase across the primary was 16 mg/l (12).

Considerable sludge thickening occurred in the primary clarifiers. Concentration of underflow sludge varied with a number of factors, including the nature of the raw wastewater. When a nearby cannery was in operation, underflow solids concentration did not exceed 4.5 percent when operating at pH 11. This seemed due to anaerobic action and gasification of peach and tomato solids in the raw sludge. Otherwise, 5 percent total solids in the underflow was maintained. Occasionally underflow at total solids levels of 9 percent

Table 5-6

LIME TREATMENT RESULTS AT CCCSD (12)

	Flow: 1.30 mgd pH: 11.5 $Ca(OH)_2$: 500 mg/l					Flow: 1.12 mgd pH: 11.0 $Ca(OH)_2$: 400 mg/l				
	Influent	Control Primary		Chemical Primary		Influent	Control Primary		Chemical Primary	
	mg/l	mg/l	% Removal	mg/l	% Removal	mg/l	mg/l	% Removal	mg/l	% Removal
Total Phosphorus	9.4	–	–	0.96	90	9.2		–	2.3	75
BOD	190	103	46	50	74	192	121	37	60	69
SS	199	57	71	41	79	195	57	71	47	76
TOC	107	59	45	37	65	118	68	42	48	59
Turbidity (JTU)	–	35		16		–	40		26	–
Calcium Hardness (as $CaCO_3$)	76	–	–	168	–	76			156	
Magnesium Hardness (as $CaCO_3$)	96	–	–	40		103			59	

Table 5-7

LIME AND METAL TREATMENT RESULTS AT CCCSD (12)

	Flow: 1.19 mgd pH: 11.0 $Ca(OH)_2$: 400 mg/l $FeCl_3$: 14 mg/l					Flow: 1.20 mgd pH: 10.2 $Ca(OH)_2$: 289 mg/l $FeCl_3$: 24 mg/l				
	Influent	Control Primary		Chemical Primary		Influent	Control Primary		Chemical Primary	
	mg/l	mg/l	% Removal	mg/l	% Removal	mg/l	mg/l	% Removal	mg/l	% Removal
Total Phosphorus	9.5	–		0.85	91	9.4	–		0.68	93
BOD	210	109	48	53	75	178	106	40	59	66
SS	305	69	78	27	91	235	59	75	31	87
TOC	130	72	45	37	72	117	68	42	43	63
Turbidity (JTU)	–	45	–	14		–	41		15	–
Calcium Hardness (as $CaCO_3$)	76	–		139		72		–	148	
Magnesium Hardness (as $CaCO_3$)	90	–		32	–	90			70	

were obtained. At about 7 percent total solids, coning and bridging occurred in hoppers of these particular tanks and septic sludge zones developed.

Contrary to experience with high-lime operation, underflow solids at pH 10.2 could not be thickened to more than 4.5 percent. Use of an anionic polymer at 0.25 mg/l allowed thickening to 6 percent. An alternate anionic polymer did not significantly improve solids underflow and neither polymer enhanced the effluent quality.

The pattern of deposition of solids in the primary clarifier was studied. Most of the sludge was found in the first three-eighths of tank length. The heavier solids settled out rapidly in this zone and the rest of the tank served to settle finally divided solids. Continuously operating scaper flights kept the bottom relatively clean. The rectangular primary clarifier, with preaerated grit removal, seemed suitable for the lime precipitation operation.

Primary clarifier effluent flowed to the activated sludge nitrification-oxidation tanks directly with no continuous intervening recarbonation stage. External CO_2 was added to mixed liquor only when needed by vaporization form liquid storage. Normally, CO_2 was only necessary when lime addition operations were at pH 11.5 or higher. In-process CO_2 generation was sufficient for complete recarbonation with primary operation at pH 11. Supplemental CO_2 was injected to the first bay of the activated sludge tanks to avoid scaling in the transfer channel. All CO_2 sources contributed to lower pH to 7.5 in a short time.

CO_2 generated in the biological process from the oxidation of carbon provided most of the recarbonation requirements in all operating modes. The amount of CO_2 generated from oxidation was calculated from a total carbon balance around the oxidation-nitrification stage and typically amounted to 55 mg/l of CO_2 for recarbonation.

The amount of CO_2 generated from the hydrogen ions produced during nitrification was calculated from alkalinity destruction, and appeared to amount to some 115 mg/l under typical conditions. The denitrification operation, occurring in a subsequent unit, was not involved in recarbonation of primary effluent.

Daily supplemental CO_2 requirements averaged about 25 mg/l after the process was stabilized at pH 11.5. CO_2 demand varied diurnally according to the strength of the wastewater.

The CCCSD treatment facility was equipped with a high-pH alarm to warn operators of the need for exterior CO_2. The 30 mgd water reclamation plant will use a fully automated pH control system.

It was recognized that, in this particular plug-flow activated sludge system, the highest pH levels would occur in the first part of the aeration tankage. With operation in the conventional mode, primary influent and return activated sludge blended at the influent end. Thus a large part of the oxygen demand also occurred in the first part of the tank.

Enough oxygen to match the load in the head end of this tank would maximize CO_2 generation there. In the 30 mgd reclamation plant, DO probes in each pass will allow varying the diffused air input according to load.

An unusual feature of this plant was application of two-stage wet classification of raw sludge by centrifuges. The first-stage centrifuge received primary clarifier underflow (thickened in situ) at a high rate. It captured two-thirds of the solids present: the resulting cake was 50 percent solids by weight and contained 85 percent of the calcium carbonate in the feed sludge. Feed sludge was not conditioned with chemicals. This first-stage separation provided sludge for recalcination. The centrate from the first centrifuge proceeded to a second centrifuge designed for efficient capture of solids remaining in the feed. When treated with polymer a drier cake was delivered. This cake which contained most of the organics and phosphorus precipitated in the primary clarifier, was incinerated. Further information on sludge handling is presented in Section 7.5.

5.4 Other Experience

This section contains information on lime treatment plants under construction, on full-scale studies, and on related information which may be of benefit to designers of phosphorus removal facilities. This is not intended to be a complete listing, but should be representative.

5.4.1 Rochester, N.Y.–Low-Lime Process

Bench scale treatability studies for phosphorus removal preceded design of a 100 mgd treatment plant at Rochester, New York (13, 14). Activated sludge treatment alone removed about 20 percent of the influent phosphorus (8 mg/l as P). This agreed reasonably well with phosphorus required for biological metabolism.

Jar tests of chemical treatment were performed. These tests examined alternated chemical procedures to achieve required 80 percent phosphorus removal. Lime, and ferric chloride with anionic polymers, were tried on raw wastewater. Either 100 mg/l of lime as CaO (pH 9.5) or the combination of 40 mg/l $FeCl_3$ and 0.5 mg/l anionic polymer reduced phosphorus 70 percent. This removal plus biological removal raised overall reduction to 80 percent or more. BOD and SS reductions in jar tests were about 50 percent and 85 percent respectively. In this case, lime was selected due to: 1) overall economy, and 2) ease of sludge handling in the plant configuration to be described.

Existing treatment plant facilities at Rochester include primary clarifiers, Imhoff tanks, sludge holding tanks, coil filters, and sludge incinerators. Proposed improvements include modification primary clarifiers to add flocculation sections, addition of new primary clarifiers incorporating flocculation sections, conversion of Imhoff tanks to complete-mix activated sludge units, and addition of chlorination facilities. Plant startup is scheduled for late 1975. Detention in flocculation sections of existing clarifiers will be 20 minutes. In the new primary clarifiers, flocculation time will range up to 40 minutes. Existing and new primary clarifiers will have overflow rates of 1,000 gpd/sq ft and detention times of 2.3 and 2 hr, respectively, at average design flow. A schematic diagram of the expanded plant is shown on Figure 5-4. Plant design is based on the criteria shown in Table 5-8.

Lime will be added just after flow passes the comminutors. The pH will be monitored ahead of the flocculation basins and lime addition will be automatically regulated to maintain pH 9.5.

Anionic polymer will be fed just ahead of the flocculation basins when such addition is necessary to improve settling. Provisions for returning solids from the primary clarifiers to the flocculation basins are included.

Although lime treatment and biological treatment should remove 80 percent of incoming phosphorus, there will be additional facilities for adding alum to aeration tanks. This subsystem may be used alone or may supplement lime treatment for greater phosphorus removal.

Complete-mix activated sludge is expected to disperse the primary effluent rapidly, permitting recarbonation by bacterial CO_2. F/M can be controlled over a broad range in the configuration provided for aeration tanks.

Since 80 to 90 percent SS removal is expected in the primary clarifiers, the weight of primary sludge (based only on increased SS removal) will be 1.5 to 2.0 times conventional primary sludge production. Sludge produced by chemical precipitation must be added to

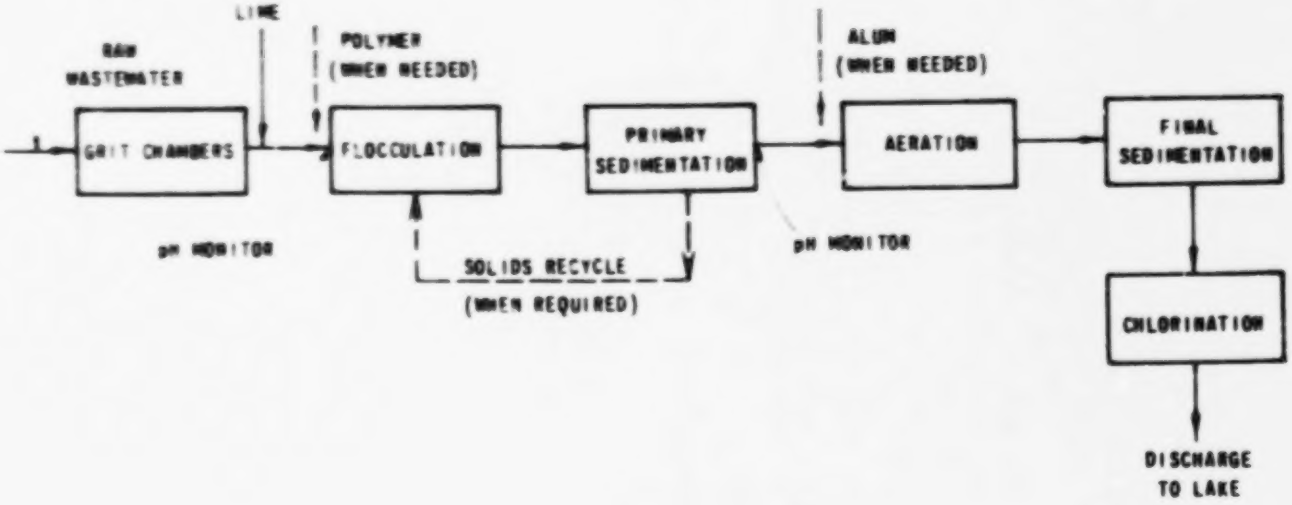
Table 5-8

DESIGN DATA FOR ROCHESTER, N. Y.

Average design flow, mgd	100
Maximum flow, mgd	200
Minimum flow, mgd	50
BOD loading, lb/day	250,000
BOD concentration, mg/l	300
Suspended solids concentration, mg/l	300
Volatile suspended solids concentration, mg/l	240
Total hardness, mg/l	200
Phosphorus, mg/l as P	8

FIGURE 5-4

TREATMENT SYSTEM AT ROCHESTER, N.Y.


RAW WASTEWATER
GRIT CHAMBERS
LIME
pH MONITOR
POLYMER (WHEN NEEDED)
FLOCCULATION
SOLIDS RECYCLE (WHEN REQUIRED)
PRIMARY SEDIMENTATION
pH MONITOR
ALUM (WHEN NEEDED)
AERATION
FINAL SEDIMENTATION
CHLORINATION
DISCHARGE TO LAKE

this, raising the total to about 3.0 times the normal amount. Primary sludge quantities at design flow with lime treatment are estimated at 175 tons/day. Since organic load on the activated sludge section will be reduced about 50 percent, activated sludge will be reduced substantially. At design loads with lime treatment, waste activated sludge is estimated at 30 tons/day, compared to about 75 tons/day if lime treatment were not used. Solids handling facilities have been designed for a capacity of 270 tons/day.

Sludges will be combined and gravity thickened to a concentration of 5 percent. Thickeners are designed for a solids loading of 13 lb/day/sq ft. After thickening, sludge will be dewatered on vacuum filters and incinerated in multiple hearth furnaces. Incinerator ash will be sluiced and pumped to two 0.5-acre lagoons, each having a water depth of 11 ft, operating in series. Effluent from lagoons will be pumped back to the head of the plant. Lagoons will probably have to be emptied every 9 to 12 months.

5.4.2 Lower Allen Township, Pa. – Low-Lime Process

Recalcination of sludge generated by lime addition in the primary is one feature of a new 6.0 mgd plant at Lower Allen Township, Pennsylvania (15). This plant, where startup began in February 1974, is a complete-mix activated sludge system with low-lime addition in the primary section.

Rough screening, degritting, and comminution precede lime injection. Flash mixing blends both recalcined lime and recycled seed sludge form the primary clarifier. Lime addition is controlled automatically within the range of pH 9.4 to pH 9.8. Mixed flow is divided between two flocculating clarifiers, each 85 ft in diameter and 12 ft in depth. These have a design average overflow rate of 700 gpd/sq ft.

Primary effluent is treated biologically in four activated sludge basins, each 45 ft. x 90 ft x 15 ft. Recarbonation reduces pH rapidly to between 7.5 and 8.5. Biological uptake is expected to complement lime precipitation in order that overall phosphorus reduction will be 80 percent. Phosphorus concentration in influent wastewater is 9.0 mg/l. Two final clarifiers are each 85 ft in diameter and 12 ft in depth.

Combined primary and waste activated sludges are blended and gravity thickened. After storage, sludge is conditioned before being fed to two parallel solid-bowl centrifuges. Dewatered cake from the centrifuges is fed to a multi-hearth dual purpose furnace. In the furnace, organic fractions are volatized while waste inert inorganic sludge is dried. The furnace is also designed to recalcine and recover quicklime for reuse at the head of the plant.

5.4.3 Manhattan, Kansas – Low-Lime Process

In Manhattan, Kansas, research into phosphorus removal using low lime and biological treatment began with laboratory scale investigations (16). It was found that both orthophosphates and polyphosphates could be precipitated readily. Complex phosphates present in wastewater seemed to inhibit calcium carbonate precipitation, to the extent that recalcination for lime recovery seemed impracticable. In these studies, 150 mg/l of slaked lime raised pH to 9.5. Total phosphorus removal before biological treatment was 60 percent, and SS were reduced 90 percent. Lime-primary sludge production was about twice that obtained in conventional operation and overall sludge production, including waste activated, was about 1.5 times that found with conventional secondary treatment.

The low lime process was operated at a constant pH. Performance of the downstream complete-mix activated sludge system was not disturbed. Microbial CO_2 production in the activated sludge unit was sufficient to maintain pH near neutral. A mixture of lime-precipitated primary sludge and waste activated sludge was dewatered and vacuum filtered with small additions of anionic polymers.

A pilot scale demonstration of the technique followed (17). The wastewater was typical domestic wastewater generated at the rate of 80 gal/capita/day. Untreated wastewater had a pH near 7, and other characteristics were BOD and SS near 200 mg/l, total phosphorus of 10 mg/l, alkalinity of 235 mg/l, hardness of 150 mg/l, calcium of 110 mg/l, and magnesium of 40 mg/l. The pilot facilities included a 15,000 gpd conventional treatment process, with lime addition to the primary. Influent wastewater was supplied to the system on a diurnal pattern from a variable speed pump and lime slurry was injected into the discharge line of that pump. Flow then entered a mixing and reaction tank where detention was near 2 minutes. Lime feed was made proportional to flow initially, but was later converted to pH control.

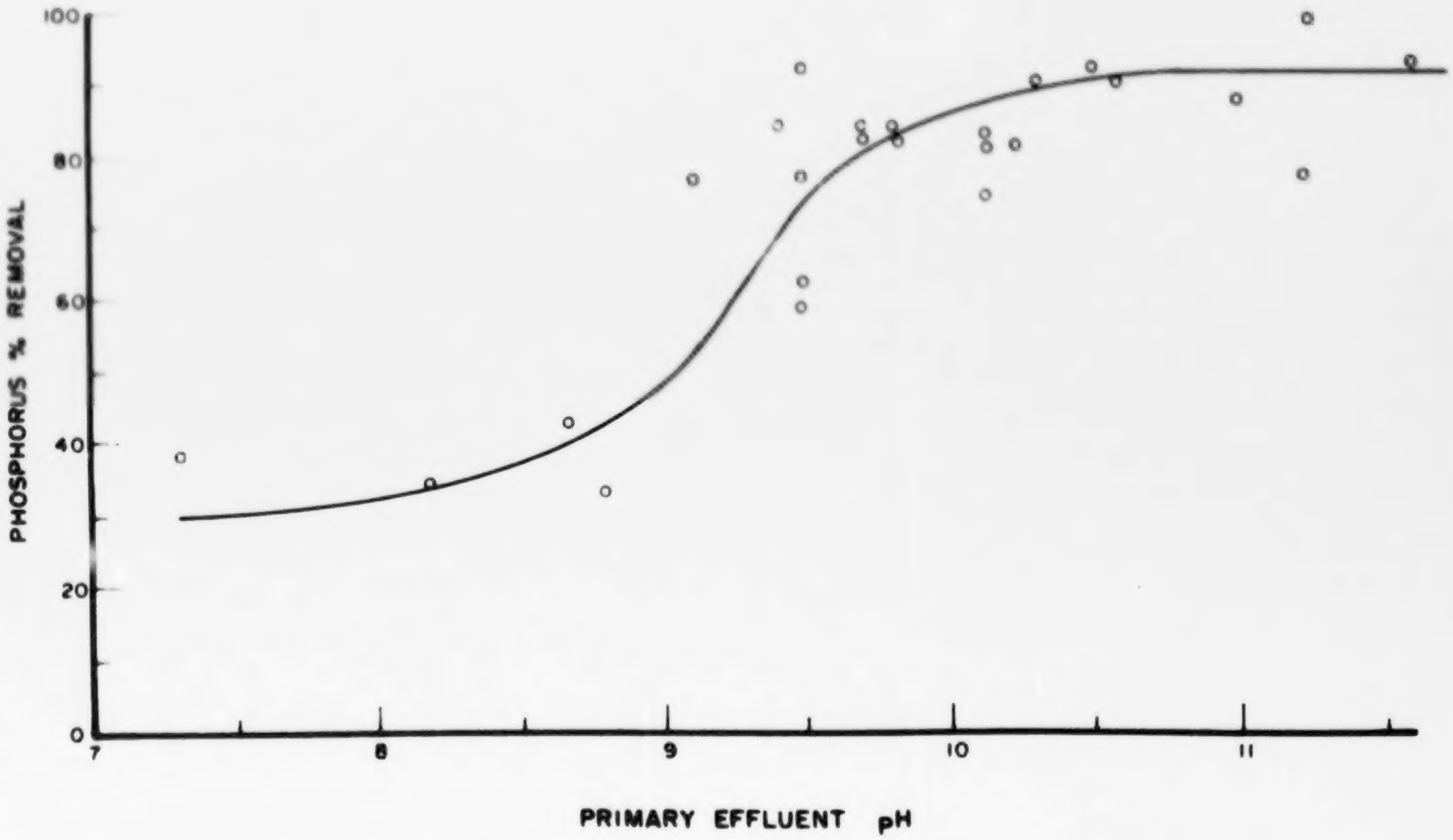
Primary clarification followed with a surface loading rate of 225 gpd/sq ft. Primary sludge was extracted hourly. Primary effluent flowed into plug flow aeration tankage where it was detained about 5 hr. The final clarifier had an overflow rate similar to the primary clarifier but sludge withdrawal was continuous.

Primary and secondary sludges were detained separately in holding tanks large enough to store the sludge from an entire day's operation. This permitted more effective sampling than could be obtained by compositing smaller portions. The two sludges were then blended into a third tank for dewatering studies. The biological system operated to give sludge age of 12 to 15 days when MLSS was 2,000 mg/l.

During the pilot study, a number of the operational features were changed deliberately or through circumstance. It appeared the most important operational variable was pH following addition of lime to the primary effluent. If that pH was held between 9.5 and 10, phosphorus removals of 75 percent or more were generally obtained in the primary unit (see Figure 5-5). Consistent phosphorus reduction was not obtained unless the pH was kept at 9.8. When pH was increased to 11.5, 90 percent phosphorus removal resulted, but a considerably greater lime dosage was required.

FIGURE 5-5

TYPICAL EFFECT OF pH ON PHOSPHORUS REMOVAL



It appeared that fine floc escaping the primary clarifier contained, in particulate form, most of the 25 percent total phosphorus residual which entered the biological section. Reduction of the flow rate (which was already quite low in terms of surface loading of the primary clarifier) did not materially improve capture of precipitated phosphorus. In this instance a higher degree of performance was not obtained by the use of polymer.

Phosphorus removal in the biological section was 5 percent or less of the total phosphorus coming into the treatment plant.

Lime treatment to pH 9.5 to 10 resulted in primary BOD reductions near 60 percent, a considerable improvement over the usual 30 or 40 percent obtained in plain clarification. SS concentrations were not materially reduced, as already mentioned in connection with lime floc which was escaping the primary unit. CO_2 production by bacteria in the activated sludge unit neutralized final effluent to levels near pH 8.

Addition of lime to the primary clarifier resulted in physical and biological changes in the activated sludge. Mixed liquor turned gold or light brown instead of the usual chocolate brown characteristic of this operation. Bacteria in the culture tended to shift into filamentous species. Among the scavenger organisms, population of free swimming ciliates declined, while rotifers increased to dominance.

The technique produced two to three times more total weight of sludge than generated during conventional treatment. However, with properly administered polymer, the sludge could be concentrated. The most attractive sludge treatment systems involved polymer conditioning, dewatering by vacuum filter or centrifuge, and incineration or land fill. Primary sludge had a volatile fraction of 50 percent during lime addition, as opposed to 65 percent in conventional systems. Chemical sludge would compact to a 3 percent density in 0.5 hr. This concentration could be considerably improved with an anionic polymer. Biological solids were produced at the rate of about 50 mg/l as the flow passed through the aeration tank. Biological sludge represented 7 percent by weight of the total sludge produced when using the lime technique.

Total sludge resulting from lime treatment was estimated at 3,750 lb/million gal of flow. This compared with 1,500 lb/million gal, of which some one-third would be waste activated sludge. Sludge conditioning with anionic polymer (2.5 lb/ton dry solids) yielded a filtration rate of 5 lb/sq ft/hr.

Several problems developed during the course of the study, and these should be considered by designers of such plants. Milk of lime solution had a marked tendency to scale everything it contacted prior to injection into the wastewater. Following injection, lime depositing out of treated wastewater caused scaling problems on the membrane of a pH probe, and on wetted parts of an automatic phosphorus analyzer. These and other surfaces also tended to collect a biological slime. Lime scale on pH probes was removed every other day by submersing them into diluted acid.

This pilot plant used 60 deg. hoppers for sludge collection in the primary clarifier. They did not work well with lime sludge which had a tendency to cling even to these sharply sloped bottoms. In due time denitrification occurred within that trapped layer and floating solids resulted. Observers noted that sludge also adhered to hopper walls in the final clarifier and became anaerobic prior to withdrawal. It appeared that this resulted in release of cellular phosphorus into the effluent.

5.4.4 Richmond Hill, Ontario–Low-Lime Process

Following laboratory and pilot scale studies, a three-month plant scale investigation of lime treatment for phosphorus removal was conducted using a 0.2 mgd section of a 2 mgd conventional activated sludge plant in Richmond Hill, Ontario (18). Lime was added before the primary clarifier on a continuous basis and no special mixing or flocculating equipment was provided.

Raw wastewater had alkalinity of 381 mg/l and total hardness of 401 mg/l. Average total influent phosphorus was 10.6 mg/l (as P). Normal plant operation reduced total phosphorus by 39 percent. BOD and SS removals in the primary clarifier were 21 and 37 percent, respectively.

Lime dosage of 175 mg/l produced a pH of 9.3. During lime addition, BOD and SS removals in the primary clarifier averaged 72 and 78 percent, respectively. Total phosphorus removal averaged 83 percent in primary treatment, and removal through the entire plant was 92 percent, resulting in a total phosphorus concentration in the secondary effluent of 0.9 mg/l.

Throughout the study, soft calcium carbonate scale accumulated on effluent weirs and troughs of the primary clarifier but was easily sluiced away using a water hose. There was no appreciable buildup on submerged surfaces within the primary clarifier.

Lime treatment did not upset the biological process in the test section. The pH of 9.3 in primary effluent was promptly reduced below 8 by CO_2 generated in the aeration tank. MLSS equilibrium was reached and maintained throughout the study. Due to improved primary performance, organic loading to the conventional (activated sludge system) was reduced and its behavior shifted to operation as an extended aeration process with no secondary sludge wastage.

Sludge precipitated in the primary clarifier was a mixture of calcium phosphate, calcium hydroxide, calcium carbonate, and coagulated organics and grit. This sludge was free flowing and considerably different from the gelatinous matter previously produced in bench scale studies. Sludge removed from the primary unit had a solids content of 6 to 8 percent. It gravity thickened readily to 10 to 12 percent solids, but was difficult to dewater without proper chemical conditioning. Biological activity in the lime-sludge was inhibited by the high lime content. The sludge proved free from noxious odors even after prolonged holding under anaerobic conditions. Supernatant from the chemical sludge, after 5 weeks of storage, contained no appreciable amount of phosphorus, either total or soluble. It was concluded that lime treatment could be readily controlled either in response to flow or on the basis of pH in a flash mixer or primary effluent.

5.4.5 Seneca Falls, New York

A combined biological-chemical precipitation process (proprietary name "PhoStrip") for phosphorus removal was applied in a short-term (30 day) study at the existing Seneca Falls, N.Y. treatment plant in 1973. This process involves modification of the conventional activated sludge process. A schematic of the Seneca Falls treatment scheme is shown in Figure 5-6.

The major modification to the activated sludge process is the incorporation of the phosphate "stripping" tank in the return sludge line. This tank permits the return sludge to become anaerobic which, apparently, induces the sludge organisms to secrete phosphate. The phosphorus enriched supernatant from the stripper is then dosed with lime and returned to the primary clarifier for sedimentation of phosphorus. Data obtained during the 30-day trial are included in Table 5-9.

Reported chemical costs for the Seneca Falls trial, based on a lime dosage of 24.1 mg/l (as CaO), were $2.01 per million gallons of raw wastewater. Additional costs would include the phosphate stripper tank, the lime feeding, storage and mixing facilities, and associated pumping facilities. At Seneca Falls, a spare clarifier was used as the phosphate stripper tank. It is claimed that total costs for the project due to chemical cost savings may result in significant savings compared to conventional mineral addition.

5.5 Sludge

Detailed information in Chapter 11 concerns amounts of sludge generated during lime addition, and offers design parameters required to select and size sludge treatment and disposal facilities. Additional information specifically related to sludge produced by lime addition in primary treatment operations is discussed in this section.

Plant scale centrifuge trials on underflow from lime treated primary units at Newmarket, Ontario, were considered successful (19). Centrifugation captured some 99 percent SS in the feed, and cake solids were near 30 percent by dry weight. The performance of the solid bowl centrifuge was such that centrate contained 700 mg/l SS and about 20 mg/l total phosphorus. Centrate was returned to the flash mixer preceding the primary clarifier. Polymer requirements were less than 1 lb/dry ton of solids. Overdosing of polymer caused foaming in the centrate.

Sludge production during low lime addition at Manhattan, Kansas (17) was 250 percent of the weight of sludge produced by conventional treatment. However, being primarily a grainy chemical sludge, this mixture of primary and waste activated sludge was much easier to handle, dewater, and dispose of than sludge produced during normal plant operation.

FIGURE 5-6

PHOSPHORUS REMOVAL PROCESS AT SENECA FALLS, N. Y.

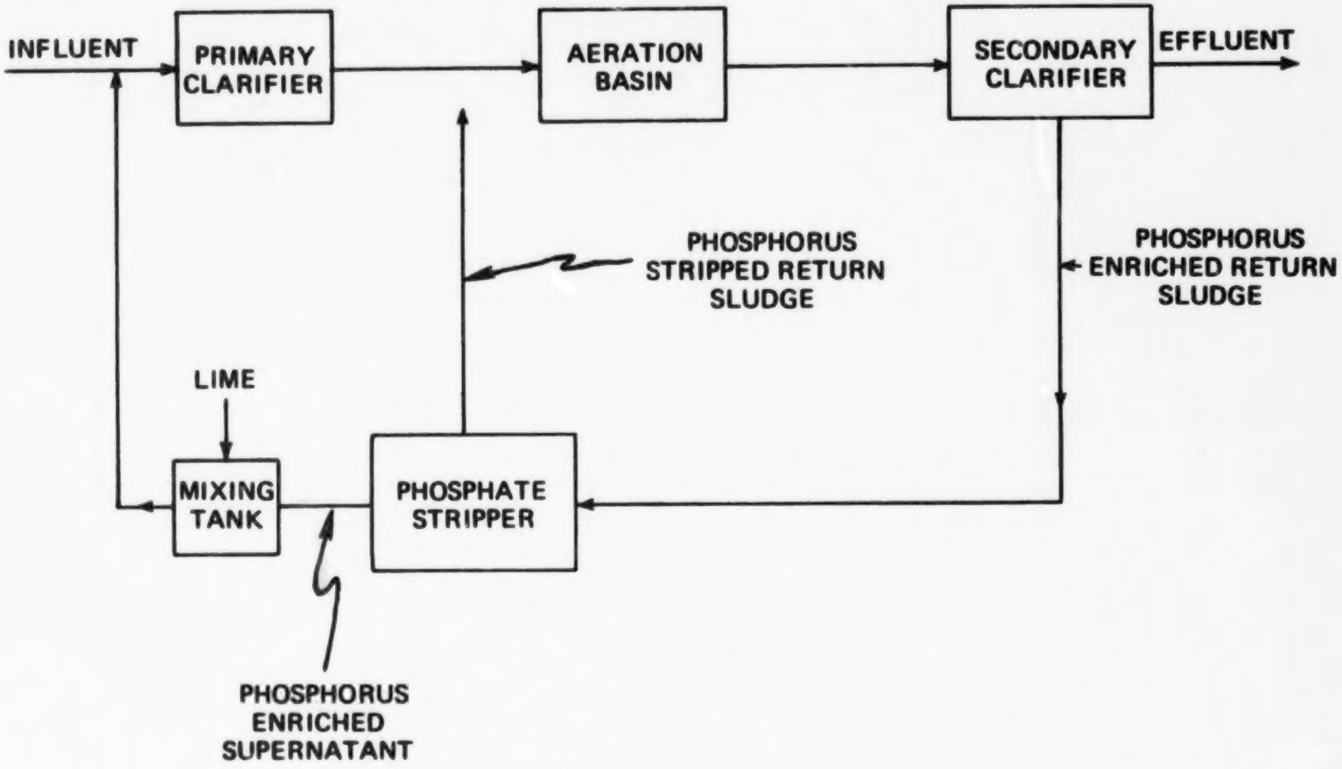

INFLUENT
PRIMARY CLARIFIER
AERATION BASIN
SECONDARY CLARIFIER
EFFLUENT
PHOSPHORUS STRIPPED RETURN SLUDGE
PHOSPHORUS ENRICHED RETURN SLUDGE
LIME
MIXING TANK
PHOSPHATE STRIPPER
PHOSPHORUS ENRICHED SUPERNATANT

Table 5-9

RESULTS OF PHOSPHORUS REMOVAL TRIAL AT SENECA FALLS, N. Y.

Plant Flow, mgd	
Design	1.75
Average	0.9
Influent	
BOD, mg/l	158
P, mg/l	6.3
Detention Times (based on average flow), hours	
Primary Clarifier	3.9
Aerator	5.5
Secondary Clarifier	3.2
Phosphate Stripper	18
Return Flows, percent of raw flow	
Sludge to Phosphate Stripper	24
Sludge to Aeration Tank	10
Supernatant to Primary Clarifier	14
Lime Dose, mg/l as CaO	24.1
SS, mg/l	
Mixed Liquor	1,440
Return Sludge to Stripper	7,840
Return Sludge to Aerator	15,910
Effluent	
BOD, mg/l	3.5
P, mg/l	0.55
Plant Performance	
BOD Removal, percent	98
P Removal, percent	91

Canadian studies (5), taken as a group, indicate total sludge production at a conventional plant will increase from 1800 lb/million gallons to 4600 lb/million gallons with addition of some 200 mg/l lime to primary clarifiers. Solids concentration increased from 4.5 to 9 percent, so the volumetric increase was 25 percent.

In a primary plant, sludge production increased from 800 lb/million gallons to 2400 lb/million gallons with solids concentration rising from 7 to 15 percent (a net 50 percent increase in sludge volume).

Plant-scale testing of two-stage centrifugation of combined lime-treated primary and waste activated sludge has been applied at Central Costa County Sanitary District (CCCSD), California (11). This permitted a degree of sludge classification vital to the integrated system of lime recovery and sludge disposal to be utilized in the 30 mgd treatment plant under construction at CCCSD. Figure 5-7 depicts the solids handling and disposal facilities which will be incorporated in the 30 mgd plant; Figure 5-8 depicts the plant-scale sludge handling system used at CCCSD to develop design data for the full-scale system.

The purpose of the first of the two centrifuges is to maximize recovery of calcium carbonate while rejecting other sludge components, such as organic material, phosphorus and magnesium compounds, and other constituents coprecipitated with calcium carbonate. The first-stage cake represents phase, all of which have a slower settling rate than calcium carbonate, are deliberately carried out with the process centrate and passed on to a second-stage centrifuge operated to obtain high solids recoveries. Second-stage solids are incinerated (breaking down organic compounds in the process) and the total ash, including fine inert materials, can be disposed of without recycling back into the wastewater treatment process.

Optimum first-stage performance in the test system was obtained after varying critical centrifuge variables: feed rate, centrifugal force, pool depth, and conveyor speed. Observations were made on individual sludge constituent recovery throughout a broad range of these process variables. Figure 5-9 shows the generalized results. The definition of performance allowed setting the centrifuge variables to capture a high percentage of calcium carbonate but a low percentage of all other sludge constituents.

Certain trends were noted in the CCCSD studies which may have application in other locations. In general, results followed past practice in centrifugation of calcium carbonate and, almost with exception, followed basic centrifuge fundamentals.

When wastewater was lime treated to a pH of 11 or 11.5, the highest percentage of calcium carbonate was recovered in the first unit and the greatest reject of extraneous solids occurred. Sludge produced when flocculating at pH 10.2 was not as easily classified. Separation was most effective when pool depth in the bowl was at very shallow settings. Dewatering of the calcium carbonate cake was materially improved by raising centrifugal force to the range of 2,100 g's. Polymer preconditioning of sludge was not required.

FIGURE 5-7 PROPOSED SLUDGE FACILITY AT CONTRA COSTA COUNTY

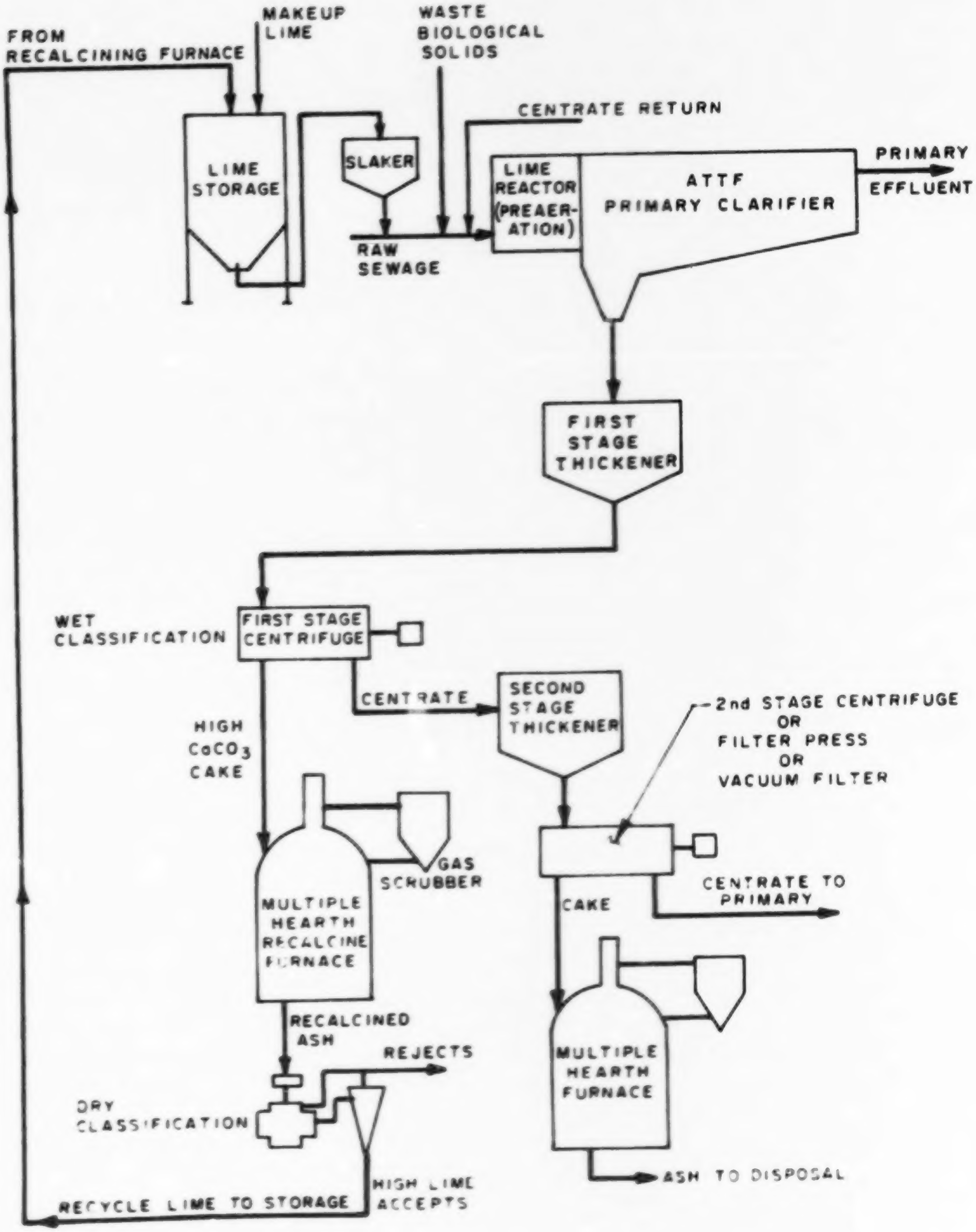
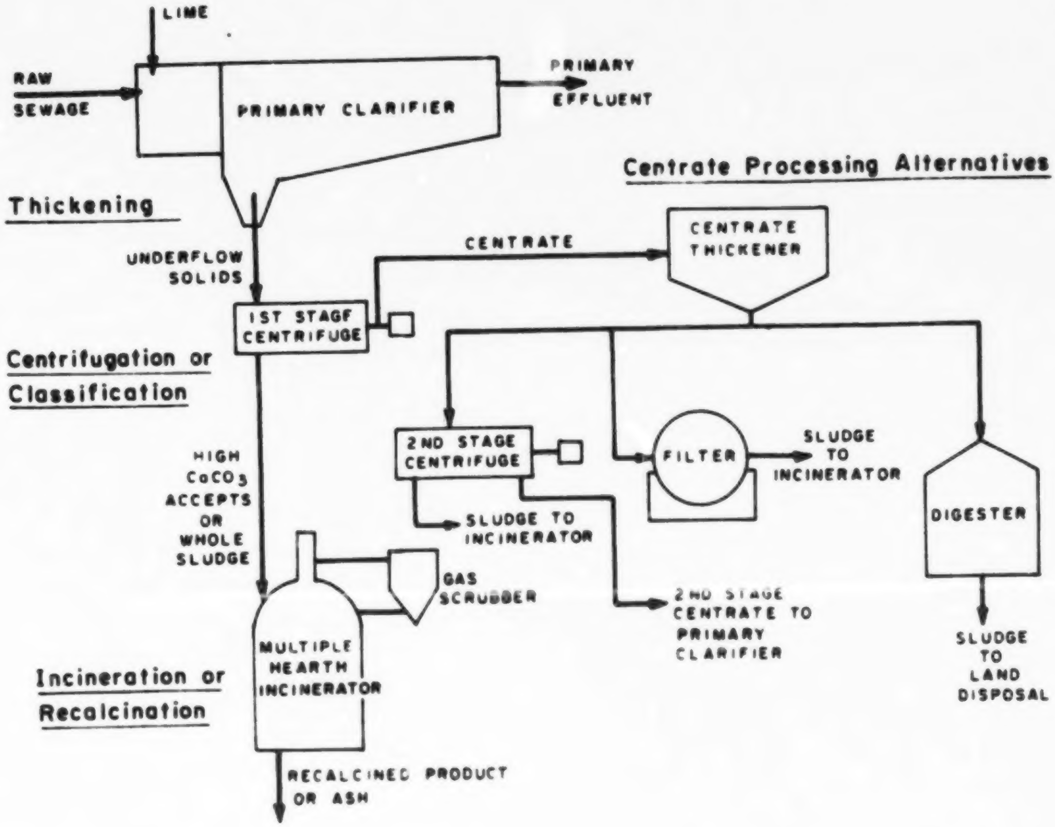
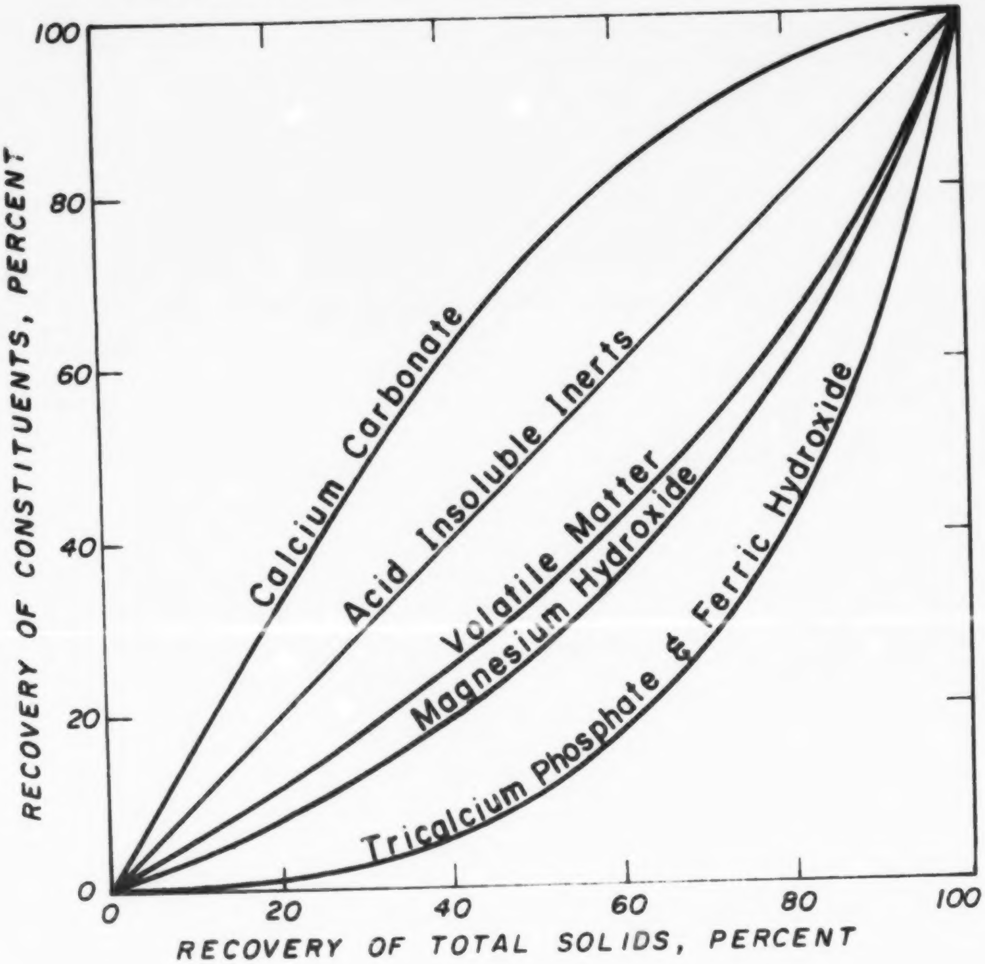

FROM RECALCINING FURNACE
MAKEUP LIME
WASTE BIOLOGICAL SOLIDS
CENTRATE RETURN
LIME STORAGE
SLAKER
LIME REACTOR (PREAERATION)
ATTF PRIMARY CLARIFIER
PRIMARY EFFLUENT
RAW SEWAGE
FIRST STAGE THICKENER
WET CLASSIFICATION
FIRST STAGE CENTRIFUGE
CENTRATE
SECOND STAGE THICKENER
2nd STAGE CENTRIFUGE OR FILTER PRESS OR VACUUM FILTER
HIGH $CaCO_3$ CAKE
GAS SCRUBBER
MULTIPLE HEARTH RECALCINE FURNACE
CAKE
CENTRATE TO PRIMARY
RECALCINED ASH
REJECTS
MULTIPLE HEARTH FURNACE
DRY CLASSIFICATION
ASH TO DISPOSAL
HIGH LIME ACCEPTS
RECYCLE LIME TO STORAGE

FIGURE 5-8 PLANT SCALE SOLIDS TESTING SYSTEM AT CONTRA COSTA


LIME
RAW SEWAGE
PRIMARY CLARIFIER
PRIMARY EFFLUENT
Centrate Processing Alternatives
Thickening
CENTRATE
CENTRATE THICKENER
UNDERFLOW SOLIDS
1ST STAGE CENTRIFUGE
Centrifugation or Classification
2ND STAGE CENTRIFUGE
FILTER
SLUDGE TO INCINERATOR
DIGESTER
HIGH CaCO3 ACCEPTS OR WHOLE SLUDGE
SLUDGE TO INCINERATOR
GAS SCRUBBER
2ND STAGE CENTRATE TO PRIMARY CLARIFIER
MULTIPLE HEARTH INCINERATOR
Incineration or Recalcination
SLUDGE TO LAND DISPOSAL
RECALCINED PRODUCT OR ASH

First-stage centrate passed to the second centrifuge for further treatment. Several polymers were tried and some proved effective in increasing solids capture when fed at the rate of 2 lb/ton of solids. Solids recovery of 80 percent and resulting cake dryness of 15 percent solids were typical operating levels obtained after experimentation. Setting the machine for maximum pool depth and operating near 2,100 g's gave satisfactory results in this particular installation.

FIGURE 5-9 SLUDGE CLASSIFICATION CURVES AT CONTRA COSTA



The extensive investigations at this installation (11) have only been briefly summarized here. The conclusion is that, in this instance, two-stage classification by centrifuges was applicable. Results of the operation were generally predictable by centrifuge laws. Other locations considering this type of operation should make comparable predesign studies.

It is also worthwhile to note that sludge production at the CCCSD test installation was predicted by theoretical procedures similar to those described in Chapter 9. Actual sludge production was then measured under four different operating modes, and operating experience agreed within about 10 percent with predicted values. These predictions and the actual production rates both included biological sludge from the nitrification-denitrification system, as well as settleable solids normally removed in primary treatment. In this case, the organic fraction of total sludge was one-quarter to one-third of the combined weight involved.

5.6 Costs

The capital cost of adding rapid mix and flocculation facilities to existing plants has been estimated for 1, 10, and 100 mgd plants. These should serve as a general guideline where phosphorus is to be removed by lime addition to the primary clarifier. Estimated costs for rapid mix facilities include concrete basins (with detention of one minute at design flow), and mixers with shafts and impellers of stainless steel. Flocculation facilities consist of baffle walls and mixers to provide a flocculation chamber with 30 minute detention within existing circular clarifiers. Estimated costs of these are given in Table 5-10 and include allowance for Contractor's installation, overhead, and profit, plus 20 percent for engineering and contingencies. Figures given relate to an EPA-STP Cost Index of 185.

Table 5-10

TYPICAL CAPITAL COSTS FOR LIME TREATMENT IN PRIMARY FACILITIES

	1 mgd	10 mgd	100 mgd
Rapid Mix	$ 2,200	$ 13,000	$ 82,000
Flocculation	64,000	264,000	983,000
Total	$66,200	$277,000	$1,065,000

Investment figures for lime storage and feed facilities are given in Chapter 10.

The cost/1000 gal for adding single-stage lime treatment to an existing plant was determined by amortizing capital cost at 6 percent over 25 years and adding the cost of chemicals. Operational expenses such as labor and power are not included.

Lime costs/1000 gal are based on a chemical cost of $22/ton with 90 percent available CaO at a dosage of 150 mg/l as CaO. The equipment and chemical costs (¢/1000 gal) are given in Table 5-11.

Table 5-11

TYPICAL TOTAL COSTS FOR LIME TREATMENT IN PRIMARY FACILITIES

	1 mgd	10 mgd	100 mgd
Equipment Amortization			
Rapid mix and flocculator	1.4	0.6	0.2
Chemical storage and feed system	0.4	0.2	0.1
Chemical Cost	1.5	1.5	1.5
Total Cost (¢/1,000 gal)	3.3	2.3	1.8

Costs do not include allowance for equipment for sludge handling and disposal and corresponding operating costs. Since a considerable quantity of additional sludge is produced, this expense is likely to be substantial.

5.7 References

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Chapter 6

PHOSPHORUS REMOVAL IN TRICKLING FILTERS BY MINERAL ADDITION

A logical approach to removal of phosphorus in trickling filter plants would be the use of precipitating chemicals somewhere in the system. Iron and aluminum can both be used for this application. Because of their similar action, iron and aluminum salts are commonly referred to collectively as minerals in wastewater treatment and are considered separately from calcium (lime). Required physical facilities for adding iron and aluminum are relatively simple. They are covered, along with special considerations for specific chemicals, elsewhere in this manual. This discussion will emphasize the merits of alternative addition points, review methods for determining dosage, and report studies where chemicals have been dosed directly to a trickling filter and to trickling filter effluent. Broader aspects of mineral addition for phosphorus removal are covered in other sections of this manual and in several recent reports (1, 2, 3, 4).

6.1 Pre-Design Decisions

State quality standards on phosphorus usually form the major basis for considering mineral addition for a given trickling filter plant. However, phosphorus concentration in the effluent is only one factor involved; effluent BOD and suspended solids will also be markedly reduced by this form of chemical treatment.

Modification of existing trickling filter plants for phosphorus removal is often a simple operation, and inclusion of chemical treatment facilities in new plants is a relatively minor addition. In either case, capital costs for chemical treatment equipment are relatively low. Also, the required degree of phosphorus removal has little effect on capital cost. Facilities for 80% phosphorus removal are identical to those for 95% removal.

On the other hand, operating costs are substantially higher to obtain successful treatment with chemical addition. More constant surveillance and more laboratory support are needed than with a conventional plant. Added operating costs are on the order of 4¢ to 5¢ per 1,000 gal. for reducing phosphorus (as P) to 2 mg/1 with an accompanying reduction of BOD and solids to 15 mg/1. For an additional 1¢ to 2¢ per 1,000 gal., phosphorus can be reduced to 1 mg/1 or less and BOD and solids to 10 or 12 mg/1. This simultaneous reduction of phosphorus, suspended solids and BOD may be the deciding factor for adopting chemical treatment in a trickling filter plant.

6.2 Process Options

There are three locations within a trickling filter plant where chemical precipitation can be incorporated: the primary clarifier, the trickling filter, and the final clarifier. The preferred approach is to provide facilities to serve both the first and last of these but not to add chemicals directly to the trickling filter.

Chemical addition to primary clarifiers will produce the greatest sludge yield of any one arrangement. Since treatment at this point will reduce influent BOD about 50%, it may be an appealing approach in an existing plant which is organically overloaded. Unfortunately, a significant amount of the phosphorus present may not be in the ortho form at this stage and may not be effectively precipitated and removed. Even after chemical treatment, primary effluent may remain fairly turbid, but will clear up during passage through the rest of the plant.

Mineral addition in the trickling filter does not seem to cause any functional problems although the filter may blotch and slough more than usual. A modest inorganic film often forms over the surface of the rocks but does not seem to interfere with biological activity or cause ponding. As data shown later will indicate, however, high degrees of phosphorus removal are not attained when chemicals are added only to the filter. Poor removal may result partly from short circuiting in the filter. This approach is not recommended.

Treatment in the final clarifier alone is a risky approach. If the operation is not done effectively, poorly treated effluent can escape to the receiving water. In spite of these reservations, experience has shown that precipitation in the final clarifier can be both effective and controllable. Orthophosphate predominates at this point and it precipitates readily. Also, bio-degradable detergents which can interfere with precipitation are largely absent. If underflow solids from a dosed final clarifier are returned to the primary clarifier, it will stimulate unusually effective clarification there.

In order to allow for incorporating the advantages of chemical treatment in both the primary and secondary clarifiers, it is recommended that chemical addition and mixing facilities be provided at both locations. Facilities required are not expensive and provision for mineral addition to both primary and secondary clarifier gives considerable flexibility of operation. Chemical treatment in both clarifiers can then be used if experience shows it to be the most effective technique at the particular plant involved.

6.3 Performance Data Using Aluminum Salts

6.3.1 Richardson, Texas

A plant-scale study was begun at Richardson, Texas, in 1970 to evaluate the potential of chemical addition to remove phosphorus from and improve overall performance of the City's trickling filter plant (5). Major objectives were to reduce the effluent phosphorus concentration to 1.0 mg/l (as P), or less, and to reduce effluent BOD and SS residuals to 15 mg/l or less.

The existing plant was a low-rate trickling filter facility with combination primary clarifier/digester units (clarigesters) as shown on Figure 6-1. Operating data for a control period before the addition of chemicals are shown in Table 6-1.

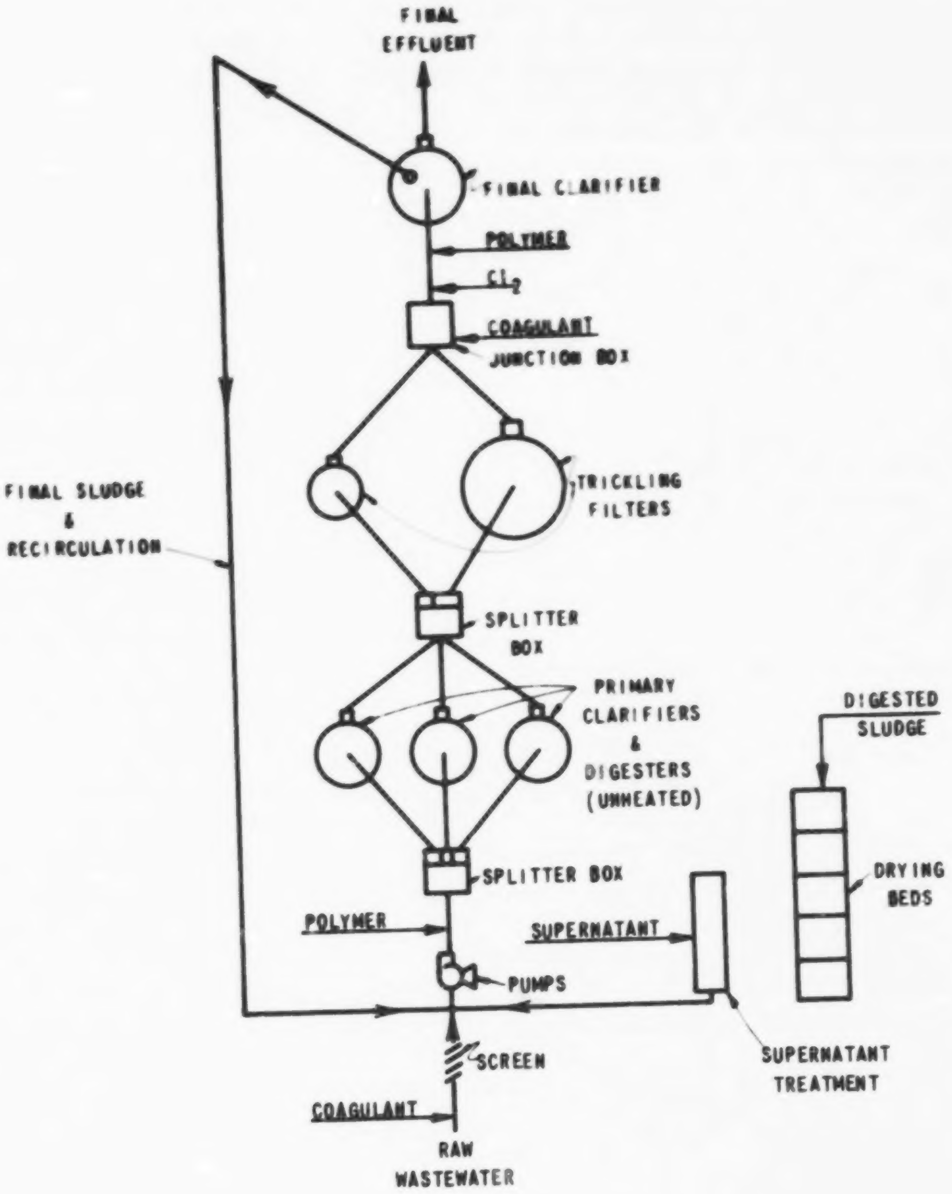

FINAL EFFLUENT
FINAL CLARIFIER
POLYMER
Cl_2
COAGULANT
JUNCTION BOX
TRICKLING FILTERS
FINAL SLUDGE & RECIRCULATION
SPLITTER BOX
PRIMARY CLARIFIERS & DIGESTERS (UNHEATED)
DIGESTED SLUDGE
DRYING BEDS
SPLITTER BOX
POLYMER
SUPERNATANT
PUMPS
SCREEN
SUPERNATANT TREATMENT
COAGULANT
RAW WASTEWATER

FIGURE 6-1 RICHARDSON, TEXAS TREATMENT PLANT FACILITIES

Table 6-1

OPERATING DATA FOR RICHARDSON, TEXAS

Description	1970 Control Period	1971-72 Extended Alum Run (11½ Months)
Flow, mgd	1.5	1.6
Influent BOD, mg/l	166	170
Influent SS, mg/l	155	155
Influent P, mg/l	11	11.4
Primary Clarifier		
Overflow Rate, gpd/sq ft	400	425
Primary Effluent BOD, mg/l	*	115
Primary Effluent SS, mg/l	*	110
Primary Effluent P, mg/l	*	8.6
BOD Removal, percent	*	32
SS Removal, percent	*	29
P Removal, percent	*	25
Trickling Filter		
Depth, ft	6.5	6.5
Recirculation Ratio	0	0
Hydraulic Loading, mgd/acre	3.9	4.1
Organic Loading, lb BOD/1,000 cu ft/day	*	14.0
Secondary Clarifier		
Alum Dosage Rate, Average Al:P mole ratio	-	1.6**
Overflow Rate, gpd/sq ft	390	415
Secondary BOD Removal, percent	*	96
Secondary SS Removal, percent	*	94
Secondary P Removal, percent	*	94
Overall Plant Performance		
Effluent BOD, mg/l	20	5
Effluent SS, mg/l	15	7
Effluent P, mg/l	8	0.5
BOD Removal, percent	88	97
SS Removal, percent	90	95
P Removal, percent	27	96

*Not reported.

**Based on Influent P.

Based on operational trials, alum addition ahead of the final clarifier was selected as the most promising chemical additive system for an extended 11½-month run. The flow diagram for the upgraded system is shown on Figure 6-1.

It was found that a mole ratio (Al : Influent P) of 1.5 to 1.7 consistently yielded effluent concentrations of 5 mg/l of BOD, 7 mg/l of SS and 0.5 mg/l of total phosphorus (as P). The corresponding effluent values for this plant prior to chemical addition were 20, 15 and 8 mg/l, respectively. The improved plant performance obtained with this upgrading technique was attributed in part to the low final clarifier overflow rate, careful management of final clarifier sludge withdrawal to prevent disruption to and loss of the alum floc blanket and frequent manual adjustment of the chemical feed pump rate to match alum dosage to mass inflow of phosphorus. Alum treatment doubled the volume of anaerobically digested sludge produced. However, the digested alum/biological sludge exhibited superior drying characteristics on sand beds and could be removed in about one-half the normal time. Operating data for the upgraded plant are shown in Table 6-1.

Chemical costs were $0.05/1,000 gal of plant flow or $0.36/lb of phosphorus removed, with phosphorus removal at the 96 percent level. The 1970 capital costs associated with plant modifications for chemical addition were $65,000, allocated as follows:

New laboratory building	$21,000
Laboratory equipment and furniture	6,000
Chemical storage and feed equipment, plant piping and metering modifications	38,000
Total	$65,000

6.3.2 Chapel Hill, North Carolina

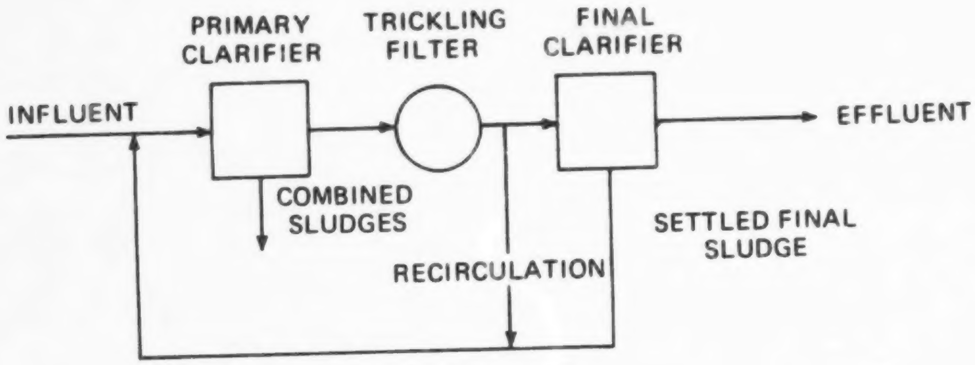
Promising results were obtained in removing phosphorus and in generally improving plant performance at Richardson, Texas, by the addition of alum to trickling filter effluent as discussed in Subsection 4.5.4.1. In view of these results, the University of North Carolina Wastewater Research Center initiated a follow-up study to further explore and confirm this process at the Chapel Hill, North Carolina Wastewater Treatment Plant (6). Conducting a similar chemical addition project at Chapel Hill was considered worthwhile because it is a typical high-rate trickling filter plant using recirculation, whereas the Richardson filters were low-rate units. Furthermore, parallel and identical lines of treatment units were available at Chapel Hill, allowing direct comparison of results with and without alum addition. Comparison of parallel results was not possible at Richardson.

The wastewater treatment plant at Chapel Hill is a conventional high-rate installation treating predominately domestic wastewater. Incoming wastewater passes through pretreatment facilities for the removal of large solids and grit. The flow is then divided into equal portions for diversion to two identical lines of treatment, each consisting of a primary clarifier, trickling filter and final clarifier as shown on Figure 6-2. On one side of the plant, a feed pump system was installed to add liquid alum to the trickling filter effluent just prior to the final clarifier.

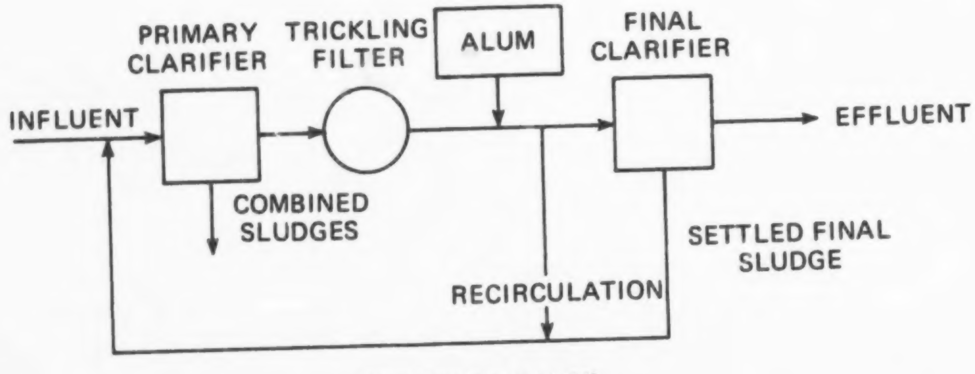
Operating data for the two sides of the plant are shown in Table 6-2. Significant improvements in BOD, SS and phosphorus removals were achieved with an alum dosage rate that ranged between an Al:Influent P mole ratio of 1.5 and 2.2. Within this range and for the loading rates shown, the final clarifier hydraulic loading appeared to be the significant factor affecting process efficiency. Subsequent phases of the study indicated that phosphorus removals of over 90 percent and BOD removals of about 95 percent could be consistently obtained by lowering the final clarifier overflow rate to 500 gpd/sq ft. Recirculation of the final clarifier settled alum/humus sludge to the primary clarifier for thickening had a beneficial effect on primary treatment efficiency and decreased the organic loading on the alum train trickling filter.

FIGURE 6-2

CHAPEL HILL, N.C. TREATMENT PLANT FACILITIES


PRIMARY CLARIFIER
TRICKLING FILTER
FINAL CLARIFIER
INFLUENT
EFFLUENT
COMBINED SLUDGES
SETTLED FINAL SLUDGE
RECIRCULATION

SIDE 1(NO ALUM)


PRIMARY CLARIFIER
TRICKLING FILTER
ALUM
FINAL CLARIFIER
INFLUENT
EFFLUENT
COMBINED SLUDGES
SETTLED FINAL SLUDGE
RECIRCULATION

SIDE 2(WITH ALUM)

Table 6-2

OPERATING DATA FOR CHAPEL HILL, NORTH CAROLINA

Description	Side 1 (No Alum)	Side 2 (With Alum)
Flow, mgd	1.41	1.41
Influent BOD, mg/l	168	168
Influent SS, mg/l	229	229
Influent P, mg/l	11.3	11.3
Primary Clarifier		
Overflow Rate, gpd/sq ft	1,100	1,100
Primary Effluent BOD, mg/l	77	63
Primary Effluent SS, mg/l	89	68
Primary Effluent P, mg/l	9.7	6.6
BOD Removal, percent	54	63
SS Removal, percent	61	70
P Removal, percent	14	42
Trickling Filter		
Depth, ft	4.25	4.25
Recirculation Ratio	2.0	2.0
Hydraulic Loading, mgd/acre[1]	16.3	16.3
Organic Loading, lb BOD/1,000 cu ft/day	18.8	15.4
Secondary Clarifier		
Alum Dosage Rate, Average Al:P mole ratio	-	1.72[2]
Overflow Rate, gpd/sq ft	885	885
Secondary BOD Removal, percent	43	76
Secondary SS Removal, percent	29	53
Secondary P Removal, percent	5	73
Overall Plant Performance		
Effluent BOD, mg/l	44	15
Effluent SS, mg/l	63	32
Effluent P, mg/l	9.2	1.8
BOD Removal, percent	74	91
SS Removal, percent	72	86
P Removal, percent	19	84

[1] Includes recirculation.

[2] Based on Influent P.

6.4 Performance Data Using Iron Salts

Addition of iron to trickling filters has been studied at Detroit and Wyoming, Michigan.

The Detroit Metro Water Department (7) compared the performance of pilot activated sludge and plastic-media trickling filter systems in order to decide on the design basis for a full-scale plant. Chemicals were added before the primary settler for both the trickling filter and activated sludge studies, and to the primary effluent for the activated sludge study. The plastic-media trickling filter test program consisted of a series of 16 experimental runs. Application rates ranged from 1.0 to about 3.5 gpm/ft^2. The raw wastewater inorganic phosphate averaged 7.9 mg/l P with a range of 1.9 mg/l to 26.6 mg/l. Without the addition of chemicals, total phosphate removals ranged from 18 to 57%. With the addition of 15 mg/l of iron (pickle liquor) removals were 34 to 77%. With 15 mg/l iron and 0.3 mg/l Dow Purifloc A23 polymer, removals ranged from 47 to 80%. With 15 mg/l iron, 0.3 mg/l A23 polymer, and 30 mg/l NaOH, removals were in the range of 67 to 78%. It was concluded that control of conditions for removal of phosphorus using the plastic-media trickling filters would be more difficult than in the activated sludge process. The activated sludge process achieved more consistent and higher phosphorus removals with iron addition and was recommended for the full-scale plant.

Dow Chemical Company performed plant scale iron addition tests at the Wyoming, Michigan trickling filter plant (8, 9, 10). The 1969 study was designed for ferric chloride addition to the raw wastewater to take advantage of the additional BOD and suspended solids removal in the primary. However, due to the recirculation of the trickling filter sludge with its high phosphorus content, it was necessary to split the $FeCl_3$ addition between the raw wastewater and the trickling filter effluent. The chemical feed and phosphorus removal data are presented in Table 6-3.

To achieve the 80% phosphorus removal required by the State, iron addition of 15 mg/l to the raw wastewater and 3 to 4 mg/l Fe^{3+} to the trickling filter effluent along with 0.3 mg/l Purifloc A23 were necessary. Removal of BOD during chemical treatment ranged fron 60% to 87% across the plant. Suspended solids removal in the primary averaged 75% with an overall removal of 87%. Iron addition to the Wyoming trickling filter plant permitted greater than 80% phosphorus removal only with the split chemical addition.

6.5 Choice of Chemical Addition

Since in-plant chemical precipitation is a new technique, pilot-scale data should be obtained prior to design of new facilities. However, when the plant already exists, full-scale trials may not be any more expensive than pilot-scale tests and they will be more reliable.

Table 6-3

IRON ADDITION FOR PHOSPHORUS REMOVAL AT WYOMING, MICHIGAN

	Test No.			
	1	2	3	4
Chemical Feed				
Fe^{3+} to raw wastewater, mg/l	15	15	15-20	15
Fe^{3+} to trickling filter effluent, mg/l	5-7	3-4	0	7
Purifloc A23, mg/l, before primary sedimentation	0.4	0.3	0.2-0.3	0
Total Phosphorus				
Raw, mg/l	10.4	11.1	14.0	17.5
Primary effluent, mg/l	1.68	2.10	3.4	6.5
Removal, %	83.9	81.0	75.7	62.9
Final effluent, mg/l	0.88	1.46	2.9	5.7
Removal, %	91.7	87.6	79.2	67.4

The jar test is a most important tool for both designer and operator when using chemical addition. Information can be obtained inexpensively and in a short period of time, however, certain precautions should be taken. An auxiliary flash mix is highly desirable if the jar test apparatus does not have capability for true high energy mixing. After intense dispersion of trial chemicals, flocculation can be simulated by careful programming of time and mixing energy levels so they approximate conditions in the plant. The most effective way to match jar test and plant conditions is to adjust the jar test mechanism until it appears to duplicate the hydraulic regime observed directly in the plant. Duration of each mixing interval should approximate plant conditions, assuming plug flow. Finally, the machine should be left turning very slowly to approximate settling conditions in the plant. A slight motion in the test jars is a better representation of clarifier conditions than if stirring paddles are turned off. No matter how carefully the rotation speed and timing intervals are selected to approximate plant conditions, the operator should devote considerable practice to use of the apparatus before his results are accepted and used.

A wide variety of other bench scale and laboratory tests (11, 12) hold considerable promise in wastewater technology. At this time, however, none of them have proved as effective as the jar test.

6.6 Nature and Role of Chemicals Involved

Compounds mentioned here are those most likely to be selected for chemical treatment in a trickling filter plant. Of the commercially available metal salts (13), liquid forms are more convenient to handle, especially at small plants. They may have the lowest overall cost and are generally more effective than dry forms. However, it may be necessary to use dry salts because transportation of pre-mixed solutions, containing both water and salt, may be costly.

Metal salt forms include liquid ferric chloride, pickle liquor, liquid alum, and sodium aluminate. Information on properties and handling these chemicals is given in Chapter 10.

Polymers are available in a variety of forms, mostly dry powders. Since the amount of polymer required is far less than the amount of metal salt, the problems of using dry polymer compounds are not severe. Of the three categories of polymers (cationic, anionic and nonionic) there is no universal choice regarding the most effective type.

Whenever chemicals are chosen, they must perform two main functions: precipitate phosphorus, and remove all types of colloids in the water. Phosphorus precipitation is caused solely by metal salts and involves conversion of soluble orthophosphate to insoluble phosphate colloids. Following that, both metal salts and polymers serve to coagulate (destabilize) colloids. Flocculation of destabilized colloids is also brought about by both metal and polymers.

6.7 Dosage Selection and Control

Proper dosage selection and control are major keys to the success of chemical treatment because this is where both cost and performance are determined. The technique selected must be effective but also should be simple and easily interpreted by the operator.

In the metal salt system, the key parameters are the mole ratio fed and the effluent phosphorus concentration. As a basis for dosage determination, complete composite hourly and daily phosphorus analyses should be performed initially. These data will provide a guide to both total daily dose required and to daily variations in dosing rate.

Once accomplished, this task need not be repeated because subsequent refinements can be based on phosphorus concentration in the final effluent. Phosphorus observations should be made at both the primary clarifier and the final sedimentation tank if direct dosing of the filter is discounted.

A suggested procedure for setting the dose is as follows. First, express incoming phosphorus in pounds per day, then convert to lb-mole/day by dividing by the atomic weight (lb/lb-mole). Atomic weights are 31 for phosphorus, 27 for aluminum and 56 for iron. Next set the mole ratio (metal:P) at the proper value, between 1.5:1 and 2:1, and compute the daily coagulant dose. An example calculation is shown below:

If incoming phosphorus (as P) is 310 lb/day,

$$310/31 = 10 \text{ lb-moles/day}$$

If desired mole ratio (M:P) is 2:1,

$$\left(\frac{2}{1}\right)(10) = 20 \text{ lb-moles metal required}$$

Using liquid alum having 4.37% Al,

(20) (27) = 540 lb Al required

(540)/(11.1 lb/gal.) (0.0437)

= 1,100 gal. liquid alum required/day

Then take the total volume to be fed per day and convert it into dosage rates with three to five changes of rate per day. These different feed rates should be selected to match incoming phosphorus at the point of injection. Dose must be matched carefully to phosphorus demand. Feeding twice the amount required part of the time and half the amount the remainder of the time will waste chemicals and lead to poor performance. In fact, overfeeding chemicals can give as poor results as underfeeding. this stems from the tendency of high metal concentrations to drive phosphorus colloids into a highly stabilized state. Dilution of sewage during rainy weather makes little difference in the amount of metal salts required since total phosphorus, in terms of lb/day, remains about the same.

Changes in feed pump settings can be made manually according to a schedule established and posted beforetime. Several types of cam regulated feed controls are also available and are effective (14).

After establishing the initial feed program, effluent phosphorus should be monitored hourly or continuously. The plant should be kept on a given feed schedule for at least three to five days, to allow the biological reactions to adjust to any effects of chemical dosing. The effluent phosphorus data will show peak concentrations resulting from dosing improperly for a period of time. The feed schedule can then be adjusted to remove these peaks. Finally, studies of different mole ratios fed can be made and the results plotted as shown on Figure 6-3. These curves can then be used to choose the mole ratio required to meet effluent phosphorus standards.

Continuing operation of the plant should be monitored carefully and adjustments should be made in the chemical feed schedule when indicated by consistently high effluent phosphorus. Chemical feeding can be automated.

Polymer may be needed for good solids removal. Control of polymer feeding is relatively straightforward. Polymer feed rates are usually kept constant, but a reduced feed rate may be used during periods of low flow. Use of polymers may reduce metal salt demand and make metal addition more attractive. Polymers which prove effective in jar tests should be carefully evaluated in plant scale operation. Typical dosage is 1 mg/l or less. One direct way to judge polymer effectiveness is to observe effluent turbidity. Polymer feeding can also be automated.

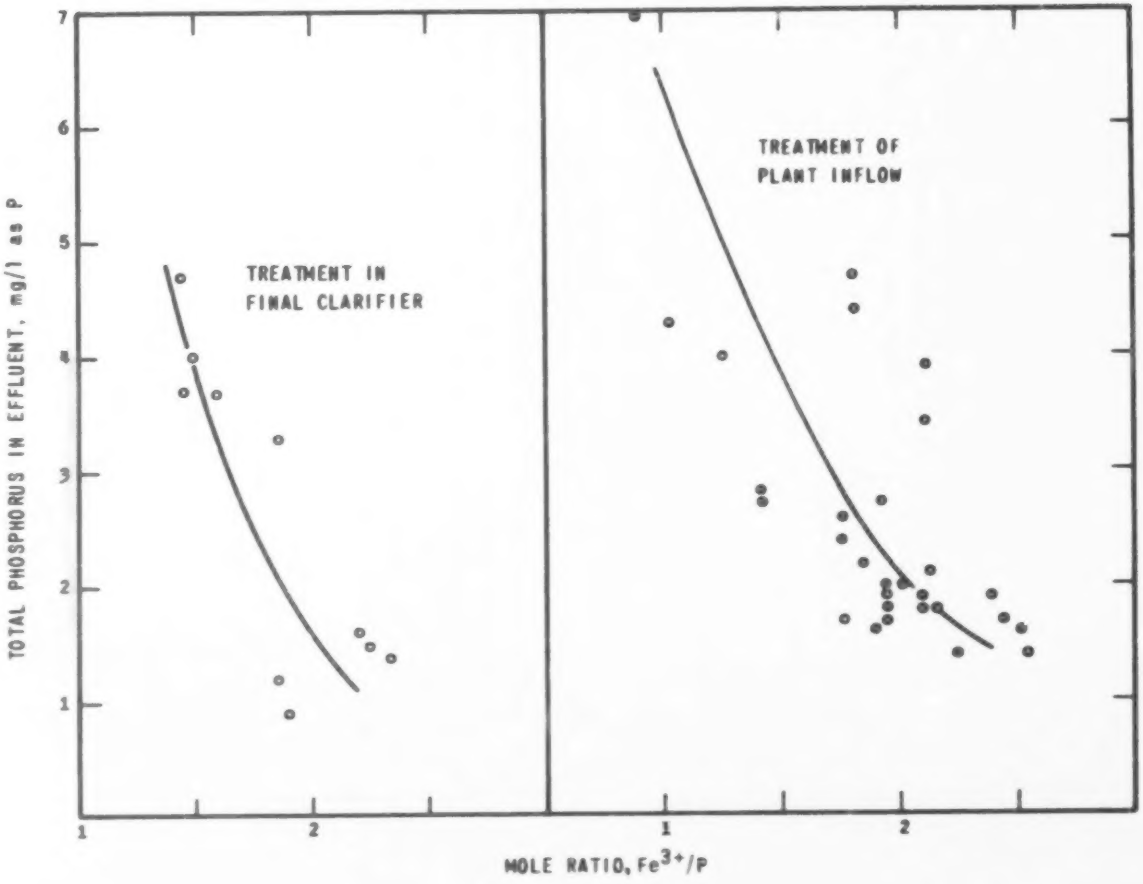


FIGURE 6-3 PHOSPHORUS REMOVAL BY IRON ADDITION (RICHARDSON, TEXAS)

6.8 Iron Leakage

Iron leakage may be a problem in trickling filter plants using iron precipitation. Figure 6-4 shows some typical data from Richardson, Texas (5). The problem appeared worse when iron was being fed in the final clarifier. It is evident that polymer treatment will reduce the escape of iron colloids.

6.9 Sampling and Analysis

The use of chemicals to remove phosphorus in trickling filter plants will require greater laboratory space, facilities, and reagents than normal treatment. Table 6-4 shows a typical matrix of data which is desirable at a conventional plant, as well as typical analyses required to support chemical treatment.

In the latter matrix, notice that the analyses include both anion and cation of the chemical being fed. Monitoring of alkalinity is optional unless the effluent concentration drops below 50 mg/l as $CaCO_3$. Some alkalinity is required to provide buffering and prevent a lowering of pH which would have an adverse effect on both chemical precipitation and biological activity.

Observation of turbidity in the final effluent is a good test. There is strong evidence that phosphorus and turbidity correlate in final effluents and the turbidity test provides a simple and responsive on-site control tool. Turbidity may be measured in the laboratory with a bench top meter or in the plant with a submersible disk lowered in the final clarifier. Observation of the height of the floc blanket in the clarifier is also a good means of control.

Effluent phosphorus concentration should be determined either on a semicontinuous basis with an automatic analyzer or on an hourly basis using automatic or manual sampling. In addition, a daily composite is also advisable. Sample portions should be flow-weighted rather than time-weighted. In plants enjoying stable effective performance, the schedule may be relaxed.

Signals from direct reading sensors and automatic analyzers can be used as a basis for automatic control of plant operation.

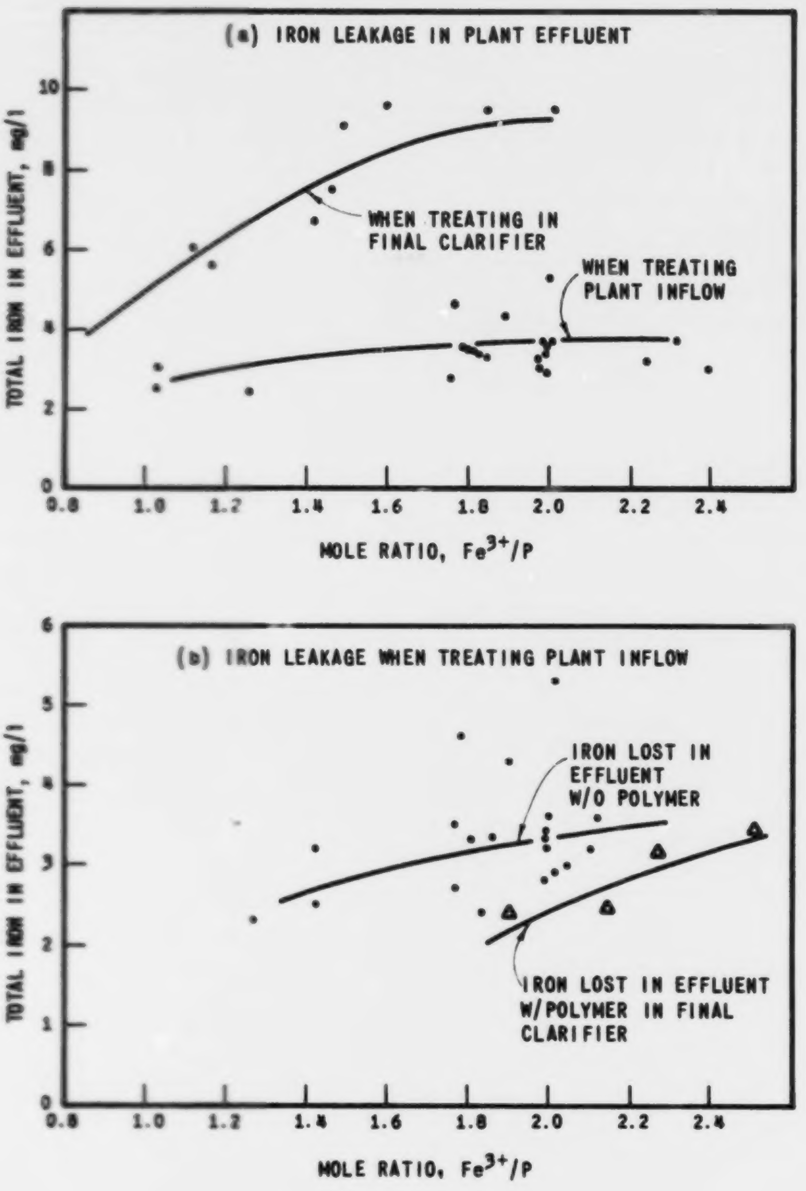


FIGURE 6-4 IRON LEAKAGE DURING PHOSPHORUS REMOVAL

TABLE 6-4 DESIRABLE LABORATORY ANALYSES

ANALYSES FOR CONVENTIONAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
FLOW	X				X	X	X
TOT SOL	X	X	X	X	X	X	X
TOT VOL SOL	X	X	X	X	X	X	X
SUS SOL	X	X	X	X	X		X
SUS VOL SOL	X	X	X	X	X		X
SET SOL	X	X	X	X			
BOD	X	X		X			X
DO	X			X			
COD	X	X	X	X	X		X
pH	X	X	X	X		X	X
TEMP	X	X	X	X		X	X

ADDITIONAL ANALYSES FOR CHEMICAL TREATMENT

	RAW	PRIM EFF	FILT EFF	FINAL EFF	RECIRC	SLUDGE	SUPERNAT
PHOS	X	X		X	X	X	X
ALK	X	X		X		X	X
FE	X	X		X	X	X	X
AL	X	X		X	X	X	X
SO_4	X	X	X	X		X	X
CL	X	X		X			
TURB	X	X		X			

6.10 References

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Chapter 7

PHOSPHORUS REMOVAL IN ACTIVATED SLUDGE PLANTS BY MINERAL ADDITION

7.1 Description of Process

The mechanism of phosphorus removal by mineral salt addition, usually iron or aluminum compounds, in a biological system is through a combination of precipitation, adsorption, exchange, and agglomeration as influenced by the pH and ionic composition of the water. The phosphorus removal technique is operationally simple and is accomplished by direct mineral addition to an aeration tank. Treatment costs are largely a function of the required effluent phosphate residual. Through optimization techniques, any degree of phosphorus removal may be provided. The main liability is the introduction of dissolved solids.

Phosphorus removal by biological systems alone, regardless of the internal mechanisms of phosphorus incorporation into the sludge solids, is limited, and removal is primarily controlled by the magnitude of solids wasting. The net amount of solids wasting is dictated by the quantity and type of substrate received and removed, system design and operation, and solids capture in the final effluent. A high rate system, with its greater solids production, would undoubtedly have the greatest background phosphorus removal potential by pure synthesis mechanisms. The magnitude of this background removal can be found by plant measurement with existing facilities or estimated by knowledge of the net solids production and phosphorus concentration in the waste activated sludge. Figure 7-1 shows the phosphorus removal capability for a variety of activated sludge systems at typically reported values for domestic wastewater treatment.

Historically, the benefits derived from mineral additions to activated sludge plants have been known since 1938 (1). Thomas (2) was among the first to point out the increased phosphorus removals. Wirts (3) also observed the increased phosphorus removals derived from a metal bearing industrial waste at a municipal treatment facility in Cleveland, Ohio. Early investigators of mineral addition for upgrading and modifying activated sludge systems for phosphorus removal include Tenny and Stumm (4), Barth and Ettinger (5), and Eberhardt and Nesbitt (6). Barth, et al., conducted pilot plant investigations with a three sludge system for combined chemical-biological nitrogen and phosphorus removal (7). Large scale, long term investigations have been recently completed at Manassas, Virginia (8), Pennsylvania State University (9), and Texas City, Texas (10). Recent publications have examined the use of aluminum (11) and iron (12) salts and investigated the kinetics and mechanisms of phosphorus removal using these minerals (13).

In the treatment of domestic wastewaters several investigators, most recently Rickert and Hunter (14), have shown that in a single sludge system a stable organic residual is found after aeration times of 30 to 60 minutes. Unfortunately, if treatment is stopped

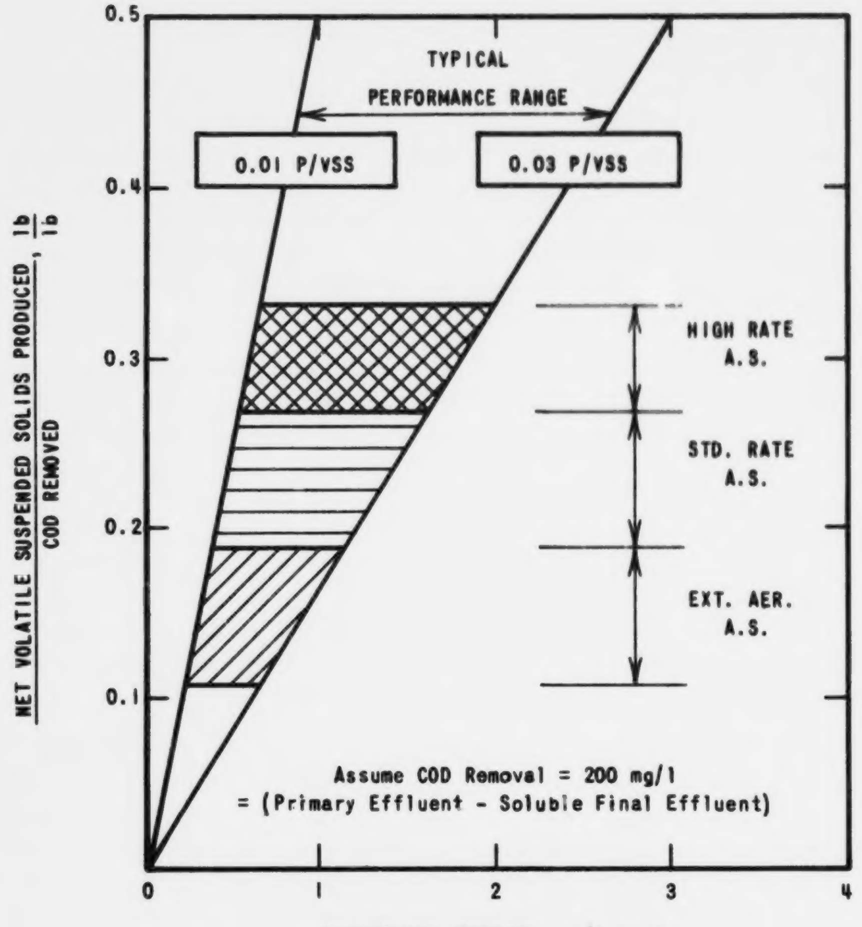

FIGURE 7-1 PHOSPHORUS REMOVAL CAPABILITY OF ACTIVATED SLUDGE SYSTEMS RECEIVING SETTLED DOMESTIC WASTEWATER

at this point, bio-flocculation is almost nonexistent and poor solids densities, settling characteristics and effluent quality result. However, when incorporating mineral addition, these liabilities are overcome and there exists an opportunity to derive capital savings from reduced tankage requirements to compensate for the increased operating expenditures associated with mineral addition. Indeed, in many cases, the use of a short residence time activated sludge system with mineral addition, settling, and filtration may result in a higher quality effluent for equal or lower costs than a purely physical-chemical approach of coagulation, settling, filtration and carbon adsorption (15). From a liability standpoint, however, the designer should be cognizant of the greater mass of generated waste activated sludge which must be successfully handled and treated for ultimate disposal.

ıneral addition to activated sludge for phosphorus removal offers the following advantages:

a. Ease of operation (direct chemical addition to existing unit processes).

b. Relatively small additional solids production (causing increases in sludge density and dewaterability).

c. Flexibility to changing conditions (treatment costs are largely a function of the influent and required final effluent phosphorus concentrations).

The main disadvantage is the introduction of dissolved solids which in some circumstances can be considered pollutants in their own right.

7.2 Mineral Selection and Addition

The selection of the mineral to use for phosphorus removal involves a multitude of considerations. Any commercial chemical, or industrial waste stream (for example, pickle liquor or alum sludge from a water treatment plant) bearing available aluminum or iron has potential application. Ideally, the best approach requires comparative full-scale plant tests of several months duration. This, of course, is usually impossible, and the next best approach is to run comparative jar tests, the results of which may not be directly applicable. However, they are at least indicative of performance. As an illustrative example, the procedures and results from the Manassas studies (8, 16) are utilized in the subsequent paragraphs.

Figure 7-2 shows the total soluble phosphorus as a function of the metal ion dose for the Manassas raw wastewater, for mixed liquor from a high rate activated sludge system, and for biologically stabilized final effluent. Since both the alum and ferric chloride can cause an alkalinity depletion and potential pH drop, pH values are also recorded. Superior performance for alum was found at all mass dosages with an apparent limiting residual of 1.0 mg/l as P for the raw wastewater and final effluent, and less than 0.1 mg/l as P for the aerator mixed liquor. A yellow color in the product water was noted with the higher iron dosages.

Replotting the aluminum data from Figure 7-2 in Figure 7-3 shows the influence of

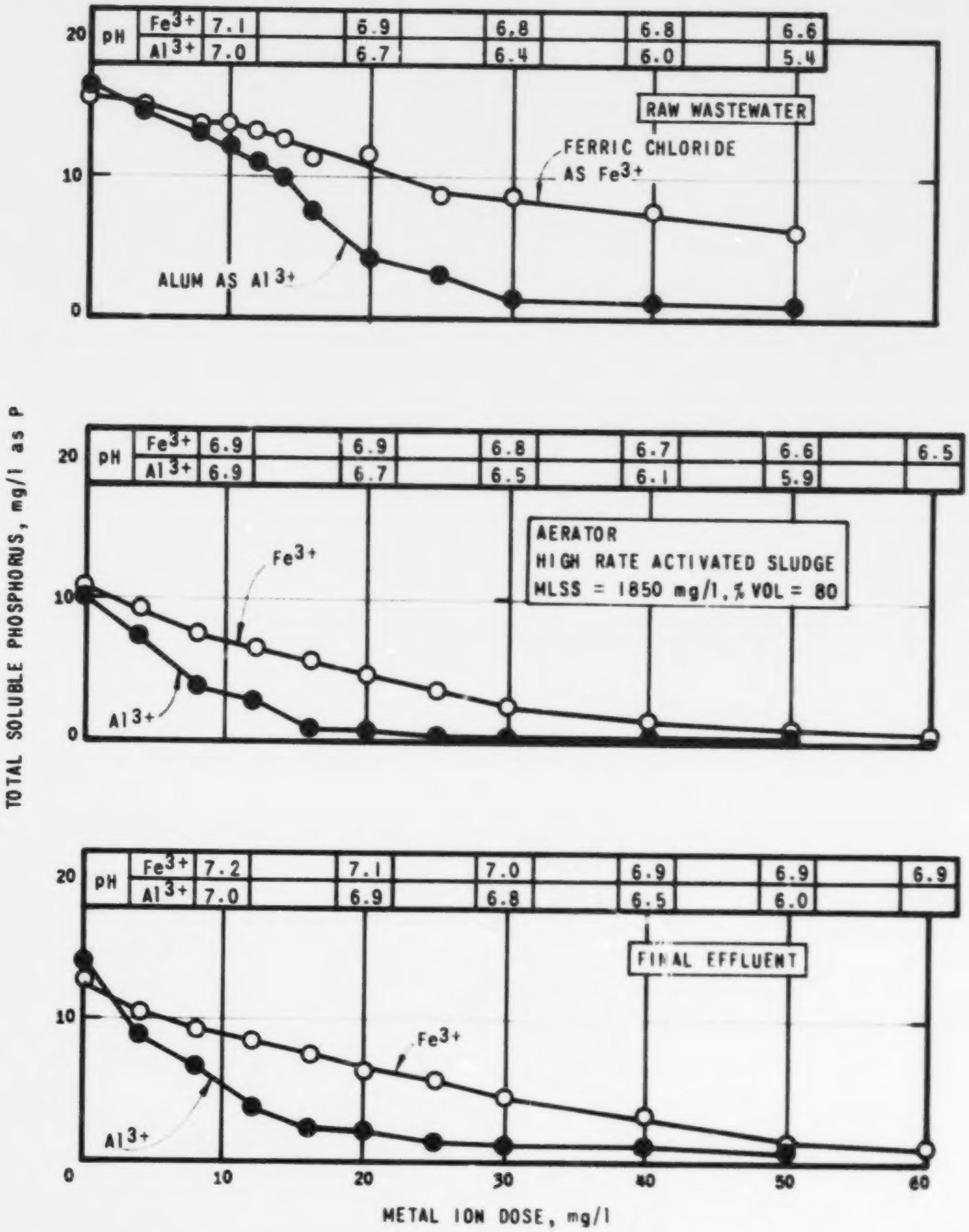

FIGURE 7-2 TOTAL SOLUBLE PHOSPHORUS RESIDUAL VERSUS METAL ION DOSE

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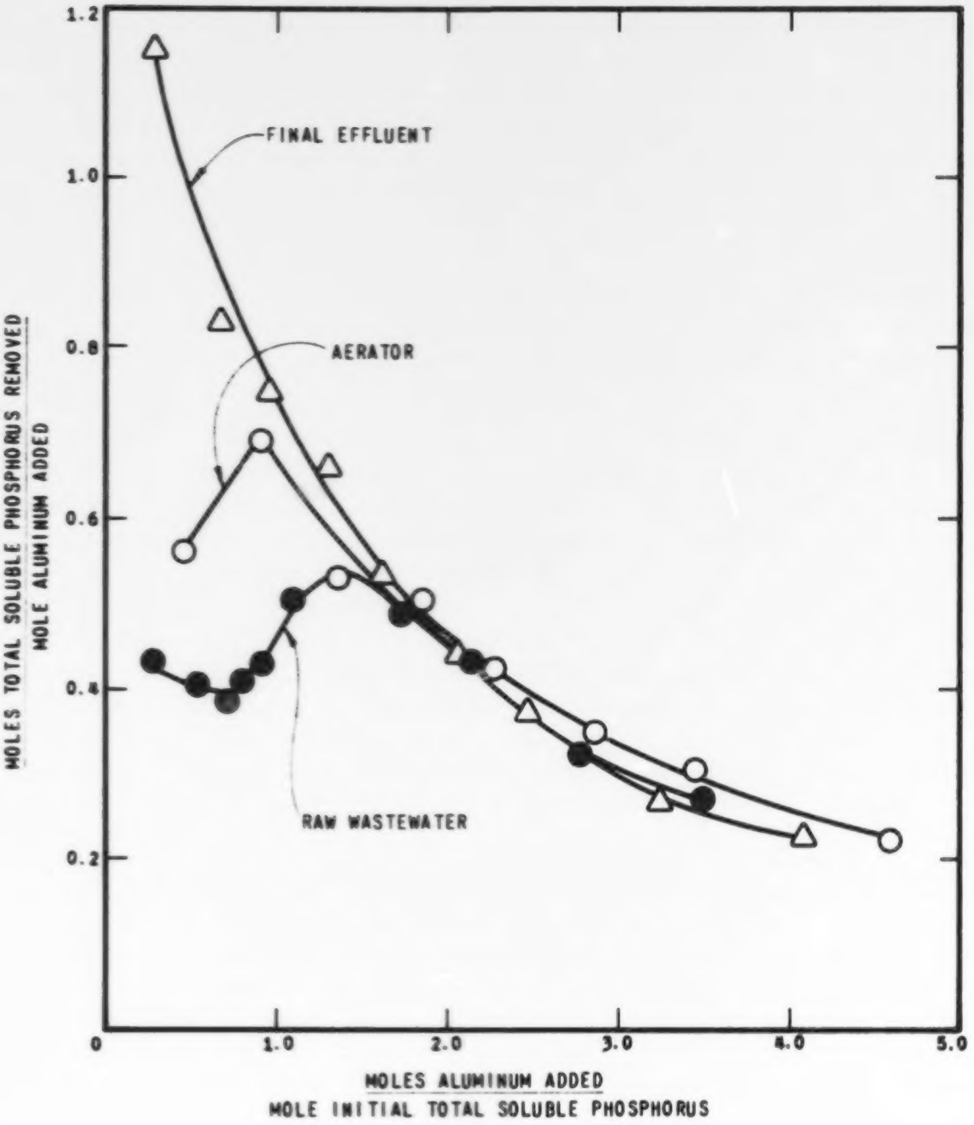

FIGURE 7-3 INFLUENCE OF POINT OF ADDITION ON PHOSPHORUS REMOVAL WITH ALUMINUM

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the point of addition on the molar effectiveness of the added aluminum. Similar but lower molar curves are found with the iron versus phosphorus residual data. The lag in phosphorus removal effectiveness encountered with the raw wastewater and aerator mixed liquor is thought to be due to the presence of unhydrolyzed phosphorus and the occurrence of competing side reactions for the metal ion. Aluminum and iron are known to form insoluble heavy salts with certain hydrophilic colloids such as detergents and soluble proteins and their degradation products (17), all of which would be present in raw and partially stabilized wastewater. These curves indicate that at molar ratios of less than 1:1, there is an obvious advantage in phosphorus removal derived by adding the chemical to a well stabilized liquid, whereas at molar ratios above 1.5:1, enough metal ion has been added to satisfy the competing side reactions and all points of addition are equally effective. The data indicate that very low phosphorus residuals are possible in a combined chemical-biological system and that the ideal point of addition would be just before final solids-liquid separation to obtain maximum biological stabilization of the soluble liquid phase.

Figure 7-4 shows the influence of wastewater and process pH upon phosphorus removal effectiveness with and without mineral additions. In the upper graph of Figure 7-4, above pH 8, phosphorus removal is occurring naturally through the formation of an insoluble calcium-phosphate complex via the natural hardness of the wastewater. The lower graph of Figure 7-4 shows that phosphorus removal at any molar dosage of metal ion is actually controlled by pH dependent reactions, optimally around 6 and 5 for aluminum and iron, respectively. Since the optimum range for aerobic oxidation of carbonaceous compounds is about pH 6 to 8, it can be seen that there is an opportunity for the simultaneous optimization of carbon oxidation and phosphorus removal by aluminum addition. However, if nitrification is a treatment goal there exists a conflict and, in turn, a compromise in pH goals. Diminished nitrification performance has been reported when alum additions have depleted the wastewater's alkalinity to a point where the pH was near optimum for phosphorus removal (9).

Table 7-1 provides a further comparison of representative sources of aluminum and iron. It should be noted that dissolved solids or extraneous ions are introduced and in some circumstances these ions may be undesirable (18). Alum and $FeCl_3$ cause an alkalinity loss, and without buffering capacity in the wastewater, a substantial pH reduction may occur. This can be compensated for by adding a source of alkalinity, such as lime, or by using aluminate. Treatment performance can deteriorate dramatically if the system pH is allowed to fall much below optimum (5). The designer should be aware that periods of heavy storm water infiltration will substantially dilute the normal alkalinity of the wastewater, and if operation is carried out at a pH of 6, only 20 to 40 mg/l of residual $CaCO_3$ buffering alkalinity may remain (8). The bottom part of Table 7-1 shows a comparison of the metal salts on a mole per mole basis, assuming a mole of metal ion precipitates one mole of phosphorus. In practice this never occurs. On this basis the possible cost advantage and alkalinity advantage with iron is lost and the aluminum salts show a potential for less additional solids production.

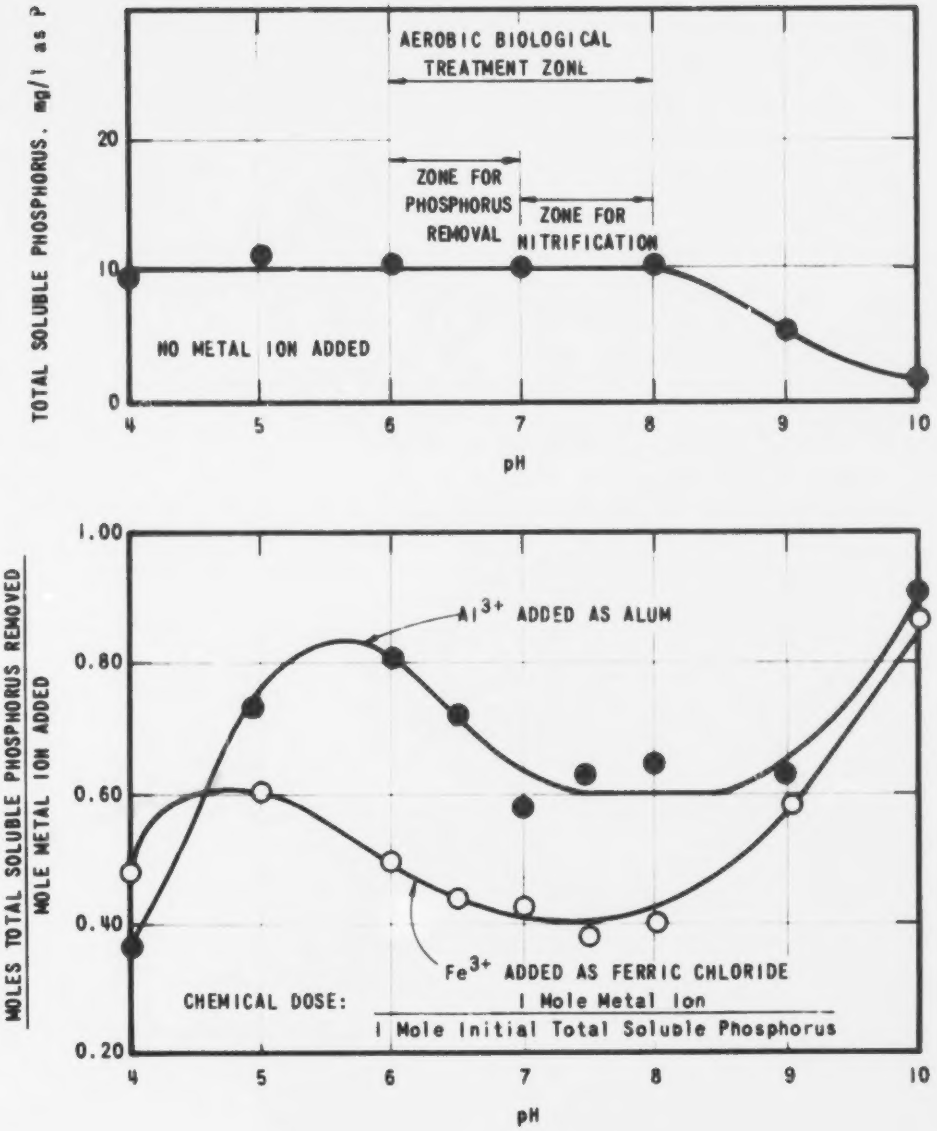

FIGURE 7-4 EFFECT OF pH ON PHOSPHORUS REMOVAL FROM A FINAL EFFLUENT

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Table 7-1

METAL SALT COMPARISON

	Metal Salt		
	Liquid Sodium Aluminate (Al^{3+})	Liquid Alum (Al^{3+})	Ferric Chloride (Fe^{3+})
	(per lb of Metal Ion)		
Bulk Delivered Price at Manassas, $	0.69	0.32	0.23
Extraneous Ion	Na^+	SO_4^{2-}	Cl^-
Introduced, lb	0.85	5.4	1.9
Maximum Alkalinity Loss as $CaCO_3$, lb	None	5.5	2.7
(Per lb of Phosphorus at a Mole Dosage of 1:1)			
Bulk Delivered Price at Manassas, $	0.60	0.28	0.41
Extraneous Ion	Na^+	SO_4^{2-}	Cl^-
Introduced, lb	0.74	4.7	3.4
Maximum Alkalinity Loss as $CaCO_3$, lb	None	4.8	4.8
Phosphorus Removal of Metal Ion, lb	0.87	0.87	1.8
Additional Solids Production, lb	$AlPO_4$	$AlPO_4$	$FePO_4$
	3.9	3.9	4.9

Several other points should be considered in mineral selection. First, in the process effluent the soluble Al, less than 0.5 mg/l (16), is apparently much lower than found with iron additions, 6 mg/l Fe reported at Cleveland (19). Second, alum adds a divalent ion (SO_4^{2-}) which has a proven coagulation benefit (20). Third, both Mulbarger (8) and Long, et al. (9), report improved product water clarity when using alum in comparison to aluminate.

Chemicals can be added with confidence in the activated sludge process just before solids-liquid separation, assuming adequate dispersal. Studies by Allied Chemical Company (21) in a 25,000 gpd pilot plant indicated that the addition of alum at the seven-eights point in the aeration basin, along with anionic polymer in the transfer line, provided the optimum phosphorus removal, coagulation and settling. Polymer dosages were found to be optimum in the 0.2 to 0.4 mg/l range. An average Al:P weight ratio of 2.1:1 was required to reduce the average influent total phosphorus of 3.6 mg/l to 0.45 mg/l in the effluent.

It should be noted that utilization of activated sludge aeration tanks for chemical mixing and flocculation is a compromise. Typically, a diffused aeration system will have a velocity gradient of over 100 ft/second/ft and the velocity gradient of a mechanical aeration system will be much higher. Water treatment experience has shown that velocity gradients greater than 75 ft/second/ft will result in the onset of chemical floc disintegration (22). Undoubtedly, some chemical floc as well as biological floc disintegration does occur in activated sludge tanks and, because of solids recycle, is unavoidable. Floc disintegration and mineral overdose both lead to product water quality degradation. In new design situations it may be advantageous to provide a brief period of gentle solids agitation prior to solids separation to promote reflocculation of the chemical-biological solids. This can be done by gentle air agitation, a flocculation tank, or a flocculator-clarifier. The designer should also provide the capability of adding organic polymer, if needed. Nalco 676 and Atlas 2A2, slightly anionic organic polyelectrolytes, have been successfully used at dosages between 0.3 and 0.4 mg/l (8).

7.3 Performance and Optimization

7.3.1 Phosphorus Removal

At molar dosages of 1 to 2, metal ion to phosphorus, soluble phosphorus residuals of 0.3 to 3.0 mg/l as P are frequently encountered in the literature. Interpretation of some of the data is complicated by failures to establish biological equilibrium [90% equilibrium values are reached at approximately 2.2 times the operating cell retention time (16)] prior to changing variables or initiating an investigative run, and failure to consider and report pH.

Detailed data from iron additions to activated sludge systems are limited in the literature and consequently the more extensively available data on aluminum addition provide a better source of definitive design information. However, a few investigations of iron additions to secondary treatment processes have been conducted on the pilot and full plant scale.

Pilot plant studies by the Detroit Metro Water Department, in which several types of activated sludge systems were investigated, indicated that the most effective total phosphate reduction was provided by the conventional activated sludge process with iron ($FeCl_2$ derived from pickle liquor) fed to the primary effluent (23). The majority of the test runs exhibited removals varying from 75 to 85% for inorganic phosphate concentrations ranging from 1.9 to 26.6 mg/l as P in the raw wastewater. Phosphate removals by the activated sludge process alone (no iron fed) fell principally in the 55 to 70% range. It was found that, in general, the addition of iron to the conventional activated sludge process provided a treatment system capable of consistently high phosphorus reduction. The step-feed, activated sludge process, with the addition of iron or iron, polymer and caustic soda, also provided consistent total phosphorus reduction. It was concluded that the full-scale plant should be designed for the activated sludge process, arranged for both conventional and step-feed configurations, and that provision be made for phosphorus removal, utilizing injection of steel pickle liquor ($FeCl_2$).

A pilot plant (5,760 gpd) at Warren, Michigan (24) was designed to simulate the existing 36 mgd activated sludge plant, with the addition of one sand filter and one multimedia filter operated in parallel after the final clarifiers. Ferric chloride was fed to the effluent of the aeration basins to determine whether or not this method would meet the State-required 80% phosphorus removal. Ferric chloride addition at a ratio of 1.5 mg/l of Fe to 1.0 mg/l of soluble PO_4^{3-} reduced the average total phosphorus concentration from 7.0 mg/l to 1.2 mg/l in the effluent. This amounted to a reduction of 83%.

Leary, et al., of the Milwaukee Sewerage Commission (25), conducted a one-year plant-scale study of the use of waste pickle liquor ($FeSO_4$) to enhance phosphorus removal. The Milwaukee Jones Island Plant consists of a single primary facility followed by two parallel activated sludge plants. Pickle liquor was employed for phosphorus removal at the 115 mgd East plant during the test period. The 85 mgd West plant served as a control. Pickle liquor feed was equivalent to an average Fe^{2+} dosage of 9.4 mg/l. The pickle liquor was added before the aeration tank about 55 feet downstream from the point of addition of return sludge. Based on a 1970 average total phosphorus concentration of 8.2 mg/l P in the screened sewage, the East Side plant, with iron addition, accomplished 91.3% phosphorus removal while the control West plant removed 83.1%. Effluent phosphorus averaged 0.7 mg/l at the East plant and 1.4 mg/l at the control plant. The return sludge phosphorous concentration was 2.29% P in the control plant and 2.61% P in the East (test) plant. The iron content of the return sludge was increased from 1.86 to 5.08%.

The pickle liquor addition increased the East plant effluent SO_4^{2-} concentration from 123 to 145 mg/l and decreased the alkalinity from 213 to 169 mg/l as $CaCO_3$. Yearly average pH values were 7.0 and 7.1, respectively, for the East and West plants. The free acid in the pickle liquor used (two sources) ranged from 2.1 to 9.3% H_2SO_4.

Comparison of the BOD, COD, and suspended solids removal efficiencies of the West and East plants indicated that the addition of unneutralized pickle liquor did not adversely affect purification, and it did not cause problems with the plant physical facilities. The results of this one-year study indicated that the use of waste sulfuric acid pickle liquor as a source of iron for phosphorus precipitation and removal was practical and effective in maintaining low plant effluent phosphorus residuals.

In studies at the University of Missouri at Rolla (26), alum and aluminate at various times were fed to laboratory activated sludge units and compared to a control unit. The alum was more effective, as shown on Figure 7-5, than aluminate, based on the Al:P mole ratio. In the activated sludge units which were dosed with aluminum, the volatile mixed liquor suspended solids dropped to 50 to 60% of the total mixed liquor suspended solids, compared to a value of 80% in the control. The aluminate units had higher pH, but all pH values were in the range of 7.7 to 8.2.

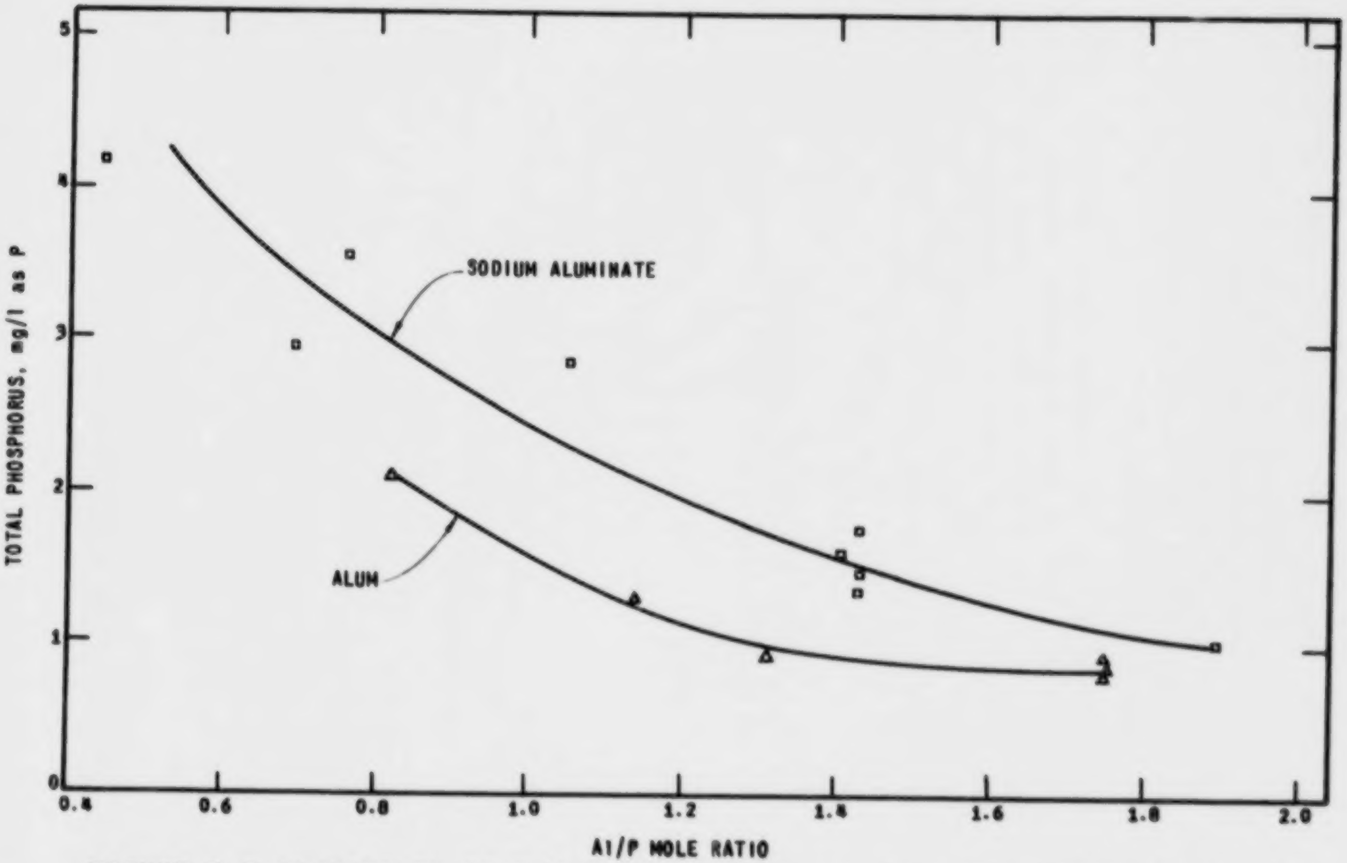

FIGURE 7-5 COMPARISON OF THE EFFECTIVENESS OF ALUM AND SODIUM ALUMINATE ON PHOSPHORUS REMOVAL

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Full-scale studies were conducted at Pennsylvania State University on the removal of phosphorus by addition of aluminum to a 2.0 mgd activated sludge plant. These studies began in 1969 and continued through 1970 (9). In the first phase of the work, alum and sodium aluminate were compared with respect to their ability to precipitate phosphorus. To prevent biological upset due to mineral solids displacing biological solids in the mixed liquor, the volatile suspended solids concentration was held at the same level. Since the volatile fraction of the total mixed liquor suspended solids decreased due to mineral solids buildup, this required the maintenance of a higher total suspended solids concentration in the mixed liquor.

The data shown by Table 7-2 indicate that alum was slightly superior to sodium aluminate in removing phosphorus from the moderately alkaline wastewater. The best point of addition for alum was found to be at the effluent channel of the aeration basin which carries mixed liquor to the final settling basin. Addition of alum both at the inlet end of the aerator and at a point about two-thirds of the way along the basin from the inlet resulted in lower removal efficiencies. Alum addition ahead of aeration produced a cloudy effluent.

Table 7-2

COMPARISON OF USE OF ALUM AND SODIUM ALUMINATE AT PENNSYLVANIA STATE UNIVERSITY

Parameter	Chemical*			
	Alum		Sodium Aluminate	
	Influent	Effluent	Influent	Effluent
Suspended Solids, mg/l	134	19	94	25
BOD_5, mg/l	87	6	50	7
COD, mg/l	244	31	160	36
Phosphorus as P, mg/l				
Filtered Ortho	6.4	0.28	5.4	0.51
Filtered Total	7.6	0.27	6.9	0.57
Unfiltered Ortho	10.4	1.17	6.2	1.14
Unfiltered Total	12.4	1.48	8.6	1.53
Alkalinity as $CaCO_3$, mg/l			156	103
pH	7.4	7.1	7.15	7.25
Alum Dosage as $Al_2(SO_4)_3 \cdot 14H_2O$, mg/l	160			
$Na_2Al_2O_4$ dosage, mg/l			44	
Al:P Ratio **, weight	2.04:1		2.04:1	
SO_4^{2-}, mg/l	24	131		

*Chemicals added at the effluent end of the aeration tank.

**Al:P ratio based on expression of phosphorus in the filtered ortho form.

Aluminate addition was found to be most effective at a point in the aeration basin near the outlet. Aluminate addition either before or after aeration produced a more turbid effluent.

Sludge production in the studies comparing alum and aluminate showed that the volume of sludge produced per million gallons of wastewater treated was about the same for both coagulants. However, the weight of solids produced was considerably less for aluminate than alum.

Because alum was found to be superior to aluminate in removing phosphorus, further study of alum was performed for a period of one year. One-half of the activated sludge plant was used in the alum study and the other half was operated without chemical addition for comparison. Table 7-3 shows average results for operation at rates equal to or less than the plant design capacity.

Table 7-3

COMPARISON OF ALUM ADDITION TO THE AERATOR AND CONVENTIONAL ACTIVATED SLUDGE AT PENNSYLVANIA STATE UNIVERSITY FOR FLOWS NOT EXCEEDING THE DESIGN CAPACITY OF THE PLANT

Parameter	Influent	Effluent with Alum Addition*	Effluent without Alum Addition
Suspended Solids, mg/l	110	22	26
BOD_5, mg/l	71	9	13
COD, mg/l	172	55	68
Phosphorus as P, mg/l			
Filtered Ortho	6.7	0.28	6.7
Filtered Total	6.3	0.36	6.7
Unfiltered Ortho	8.9	1.15	7.2
Unfiltered Total	10.0	1.41	7.3
Alkalinity as $CaCO_3$, mg/l	168	80	120
pH	7.6	6.75	7.35
Al:P Ratio**, Weight		2.97:1	

*Alum added to effluent of aeration tank. Values shown in this column are means for the period of study.

**The Al:P ratio is based on expression of phosphorus in the filtered ortho form.

Problems were experienced with loss of solids from the final basin during peak flow periods where alum treatment was being practiced. It was believed, however, that proper allowance for peak flows in design of such facilities would eliminate excessive solids loss. The effluent insoluble phosphorus concentration was found to correlate well with the effluent suspended solids concentration.

A 2,000 gpd pilot operation at the main wastewater treatment plant of the District of

Columbia was operated at a constant alum dose rate (27). The pilot plant is a four-stage step aeration process followed by secondary clarification. Alum was added to the wastewater of moderate alkalinity (100 to 150 mg/l as $CaCO_3$) at the fourth stage of aeration using an Al:P weight ratio of 2:1 (molar ratio 2.3:1), based on an influent phosphorus concentration of 8.15 mg/l as P. Mixed liquor suspended solids averaged 6,300 mg/l with a volatile content of 60% and a cell retention time of about 5 days. Total phosphorus residuals averaged 1.9 mg/l as P (78% removal) in the clarifier effluent.

Figure 7-6 shows the chemical phosphorus removal found for the Manassas high rate activated sludge plant with aluminate (16) and alum (8) additions at slightly alkaline pH values. Phosphorus concentrations for Figure 7-6 were corrected for biological removal. The Manassas Pilot Plant, shown in Figure 7-7, is a 0.2 mgd three-stage (nitrification-denitrification) activated sludge system operated at the Greater Manassas Sanitary District of Prince William County, Virginia. Table 7-4 shows the system performance for 200 determinations with alum addition. The reader is referred to the original publication for more detailed information.

Table 7-4

COMBINED CHEMICAL–HIGH RATE ACTIVATED SLUDGE SYSTEM PERFORMANCE

			Percent Less Than				
Item	Min.	10	20	50	80	90	Max.
Primary Effluent							
COD, mg/l	27	64	86	152	227	250	324
P, mg/l	2.2	3.8	5.0	8.4	11.4	12.8	18.5
BOD_5:COD*	0.26			0.43			0.56
High Rate Activated Sludge Effluent							
Soluble Phase							
COD, mg/l	13	23	26	33	41	48	79
P, mg/l	0.2	0.8	1.0	2.0	3.0	3.6	6.8
BOD_5:COD*	0.14			0.28			0.36
Solid Phase							
Suspended Solids, mg/l	0	10	14	21	36	44	196
BOD_5:SS*	0.10			0.23			0.31
COD:SS*	0.43			0.58			0.80
P:SS*	0.039			0.051			0.063
Suspended Solids After Polymer Addition,** mg/l	0	0	1	2	8	14	18
Suspended Solids After Mixed Media Filtration**, mg/l	0	0	0	0	2	5	12

*Ratio of minimum, median, and maximum monthly averages for each parameter
**94 determinations

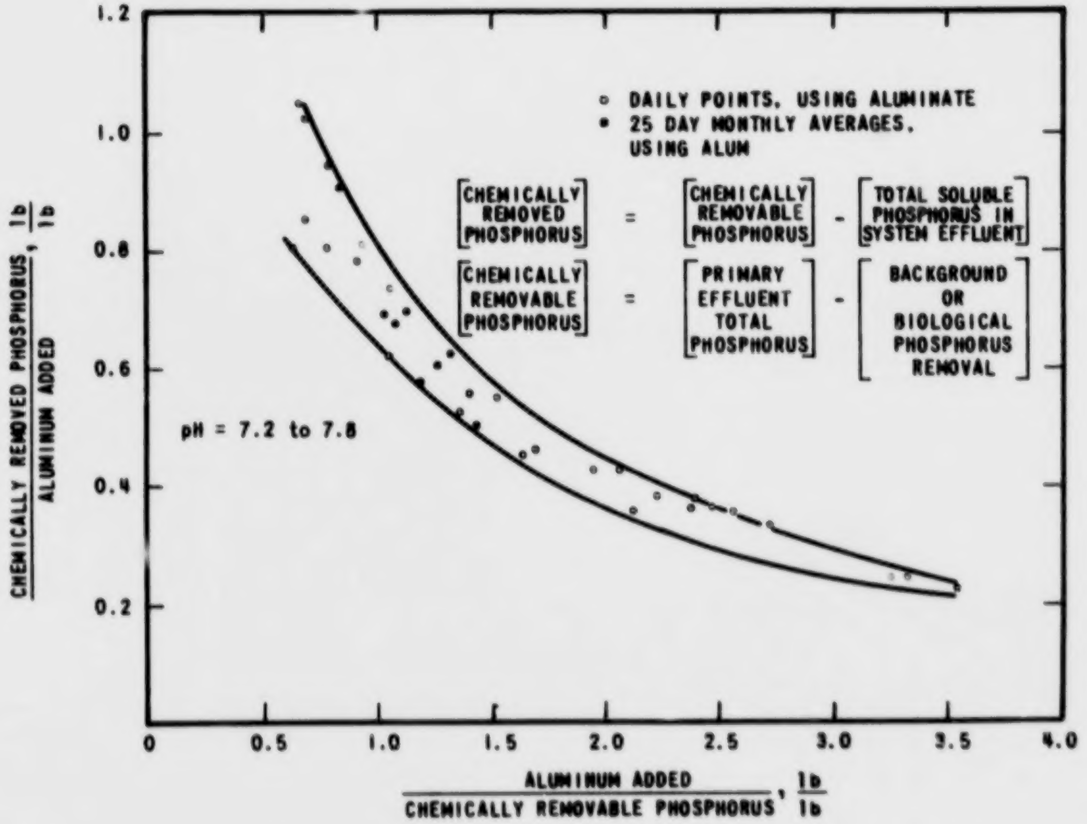

FIGURE 7-6 CHEMICAL PHOSPHORUS REMOVAL ENVELOPE: HIGH RATE ACTIVATED SLUDGE SYSTEM

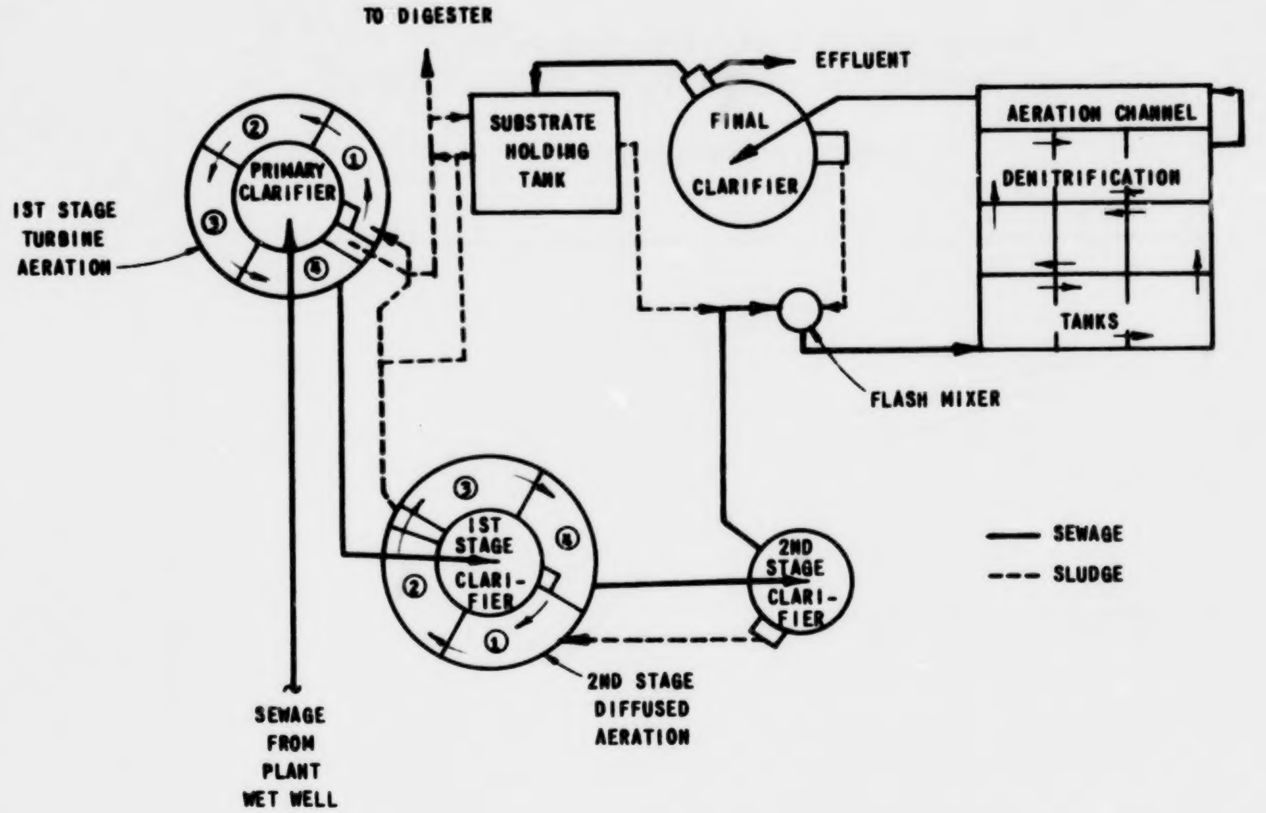
TO DIGESTER
EFFLUENT
SUBSTRATE HOLDING TANK
FINAL CLARIFIER
AERATION CHANNEL
DENITRIFICATION
TANKS
1ST STAGE TURBINE AERATION
PRIMARY CLARIFIER
①
②
③
④
FLASH MIXER
1ST STAGE CLARI-FIER
2ND STAGE CLARI-FIER
SEWAGE
SLUDGE
2ND STAGE DIFFUSED AERATION
SEWAGE FROM PLANT WET WELL

FIGURE 7-7 0.2 mgd BIOLOGICAL NITRIFICATION - DENITRIFICATION PILOT PLANT

Greater Manassas Sanitary District
Prince William County, Virginia

Optimization of phosphorus removals with mineral additions can be done two ways: by pH adjustment to the point of minimum solubility (pH 6 to 6.5, residual alkalinity of 20 to 40 mg/l as $CaCO_3$) and by splitting the chemical dose to separate parts of the treatment system (8). Optimization maximizes the efficient use of chemical, minimizes additional solids production, and produces soluble phosphorus residuals of less than 0.5 mg/l. Hais, et al. (27), indicated, using step aeration with alum addition, that above pH 6.6 a rapid increase in the effluent phosphorus residual occurred with increasing pH. However, operation with a mixed liquor pH of less than 6.2 caused an upset in the activated sludge process. It was concluded that optimum performance (0.6 mg/l effluent phosphorus residual) was achieved by maintaining a final stage pH between 6.3 and 6.6 at an Al:P weight ratio of 2:1. Mulbarger (8) showed the capability of pH adjustment with a denitrification system and demonstrated the advantages of split chemical treatment in a multi-sludge system. Results from the pH adjusted system are shown in Figure 7-8. Long, et al. (9), found a natural pH adjustment with high doses of alum in a poorly buffered wastewater and report phosphorus residuals below 0.5 mg/l at Al:P weight ratios of from 2:1 to 3:1.

Organism activity and type (9) as well as the completeness of a chemical reaction affect phosphorus removal and are influenced by the temperature. No temperature effect has been reported, however, for the rate of chemical reaction (13) operating between 10 and 20° C.

Mulbarger (8, 16) notes that the presence or absence of nonsettleable effluent suspended solids from a combined chemical-biological system is more dependent upon the ratio of net volatile solids produced to aluminum added than upon pH, exceeding an aluminum to phosphorus ratio, or exceeding a given aluminum dosage. It is recommended that this ratio not drop below 3 to 5. It is believed that the biological volatile solids provide exchange and/or sorption sites of limited capacity for the precipitated aluminum-hydroxy-phosphate complex, and that these natural polymers aid in clarification. Thus, the more biological solids produced from the system the greater the aluminum dosage that can be used without effluent suspended solids problems.

When designing the phosphorus removal system, the designer should be cognizant of the phosphorus levels in the effluent suspended solids. There was a nominal 1.0 mg/l of phosphorus in the effluent suspended solids from the high rate system at Manassas (8). After pH adjustment and addition of 0.3 to 0.4 mg/l of anionic polymer (Nalco 676), effluent suspended solids were reduced to 2 mg/l. Consequently, the effluent total phosphorus was also reduced. Polymer performance was also reported to be influenced by the pH (8). If total phosphorus residuals of less than 0.5 mg/l as P are required, a multimedia filtration system is recommended. Dual and tri-media filtration both provide effective tertiary solids separation. Hais, et al. (27) showed that tri-media filtration consistently removed between 5 and 10% more suspended material than did the dual-media filter. Filter runs between 24 and 32 hours were regularly maintained. Tri-media filtration performance was reported to be superior to dual-media filtration at Manassas (8).

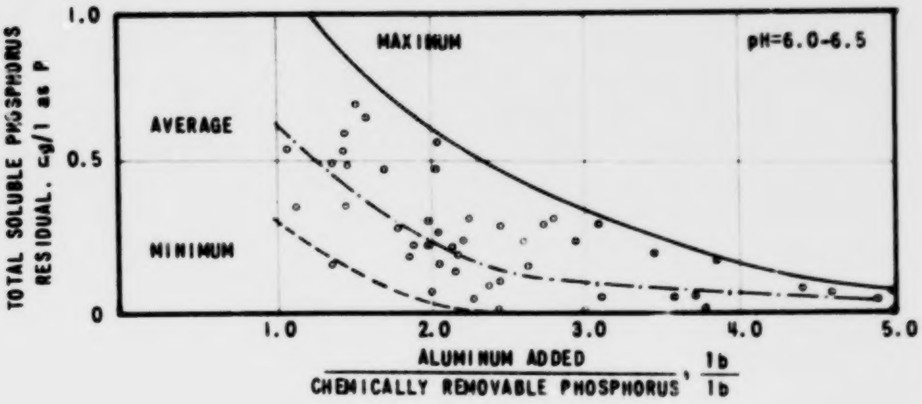

FIGURE 7-8 EFFECT OF ALUMINUM ADDITION WITH pH ADJUSTMENT ON TOTAL SOLUBLE PHOSPHORUS RESIDUAL

7.3.2 Costs

Power costs for combined chemical-biological treatment are negligible. Maintenance costs will increase somewhat due to the additional equipment. Capital expenditures for storage facilities, instrumentation, and equipment will be very low at larger facilities unless multi-media filtration is incorporated. Chemical costs will vary as a function of the incoming phosphorus concentration and the required phosphorus residual. Additional expenditures for pH control (alkalinity dependent) and solids control (polymer additions) may be necessary. See Chapter 10 for further information.

7.4 Process Design Examples

Four design example problems were selected for an arbitrary wastewater. Each illustrates specific points in combined chemical-biological wastewater treatment with alum additions for phosphorus removal. Results from the Manassas work are used for some of the calculations.

Assumptions:

a. Net biological solids production = 0.28 lb of volatile solids produced per lb of COD removed; the volatile solids contain 2% phosphorus, and the total solids average 80% volatile.

b. The biological system is to be designed with a minimum hydraulic contact time of 1 hour at the maximum 4 hour hydraulic peak. The ratio of maximum flow rate to average flow rate = 1.5. A cell retention time (CRT) of two days is selected.

c. Primary Effluent Characteristics:

$$\text{COD} = 200 \text{ mg/l}$$

$$\text{P} = 10 \text{ mg/l}$$

$$\text{Alkalinity} = 150 \text{ mg/l as } CaCO_3$$

Find:

1. Effluent COD, BOD_5 and phosphorus
2. Total solids production
3. Alkalinity loss
4. Dissolved solids addition
5. Size of final clarifier, return and waste sludge pumping systems.
6. Chemical costs with the following delivered prices:

Alum	= \$0.32/lb Al^{3+}
Polymer	= \$1.50/lb polymer
Acid	= \$0.02/lb H_2SO_4
Lime	= \$0.01/lb $Ca(OH)_2$

Example No. 1: High rate system for most efficient chemical addition with pH above 7 and no optimization by pH adjustment.

1. **Effluent COD, BOD_5, and Phosphorus**
   (See Table 7-4. Note: Soluble organics from a biological system vary as a function of the wastewater and dilution of the refractory fraction. Manassas values are used for illustration only.)

$$\text{Effluent soluble COD} = \frac{200-152}{227-152}(41-33) + 33 \quad \text{[Interpolated from data of Table 7-4]}$$

$$= 5 + 33 = 38 \text{ mg/l}$$

Assume effluent suspended solids = 20 mg/l

$$\text{Total effluent COD} = 20\,(0.58) + 38$$

$$= 12 + 38$$

$$= \underline{50 \text{ mg/l}}$$

Effluent BOD_5 = 38 (0.28) + 20 (0.23)

= 10.6 + 4.6

= 15 mg/l

[Using BOD_5 ratios from Table 7-4]

Background phosphorus removal per lb COD removed

= (biological solids production ratio) (phosphorus content)

= (0.28) (0.02)

$$= 0.0056 \ \frac{\text{lb phosphorus removed}}{\text{lb COD removed}}$$

[The background phosphorus removal rate at Manassas was found to be approximately $0.01 \ \frac{\text{lb P Removed}}{\text{lb COD removed}}$ (8)]

Background phosphorus removal expressed as concentration

= (COD removal) (phosphorus removal per lb COD removed)

= (200-38) (0.0056)

= 0.9 mg/l

Chemically removable phosphorus = 10.0 - 0.9 = 9.1 mg/l

Use Al^{3+}/chemically removable phosphorus dose = 1.25 lb/lb

From Figure 7-6 find 0.6 lb of phosphorus removed per lb of aluminum added at the dosing ratio of 1.25

Aluminum dose = 1.25 (9.1) = 11.3 mg/l Al^{3+}

Chemical phosphorus removal = 0.6 (11.3) = 6.8 mg/l P

Soluble phosphorus residual = 9.1 - 6.8 = 2.3 mg/l

Total effluent phosphorus = 2.3 + 20 (0.05)
= 3.3 mg/l P

[P/SS ratio from Table 7-4]

The preceding calculations are useful from the standpoint that they show the pollutant fractions in both the soluble and solid phases. Often a remarkable treatment improvement can result from effluent solids control and minimization.

2. **Alkalinity Loss**

Maximum possible alkalinity loss due to aluminum hydrolysis

= 5.4 (5.5) = 30 mg/l as $CaCO_3$

which gives $\frac{30}{11.3}$ = 2.7 lb $CaCO_3$ loss/lb Al^{3+}

At an aluminum added to chemically removable phosphorus ratio of 1.16, an average alkalinity loss of 2.3 lb $CaCO_3$/lb Al^{3+} was measured at Manassas (8). Like solids production values, alkalinity losses are a function of the proximity of aluminum dose to stoichiometric phosphorus requirements and the reaction pH.

3. Dissolved Solids Addition

$$SO_4^{2-} = (11.3 \text{ mg } Al^{3+}/l)\left(5.4 \frac{\text{mg } SO_4^{2-}}{\text{mg } Al^{3+}}\right) = \underline{61 \text{ mg}/l}$$

Although dissolved solids are added to the system, the designer should remember that other soluble wastewater components (i.e., organics, phosphorus, alkalinity, etc.) are being removed. Prior to the addition of sulfuric acid for pH adjustment at Manassas, no increase in dissolved solids was measured on a primary to final effluent basis because of a compensating loss of soluble wastewater components (8).

4. Final Clarifier and Pumping System

Organic solids = 57 mg/l
Inorganic solids = 43 mg/l
Total = 100 mg/l
Solids lost in effluent = 20 mg/l
Solids accumulation = 80 mg/l

CRT = 2 days, and aeration tank detention at average flow

= 1.5 (1) = 1.5 hours

$$\text{MLSS in aerator} = \frac{(24)(2)}{1.5}(80) = 2{,}560 \text{ mg/l}$$

Since this is the final clarifier a conservative overflow rate should be utilized. From Figure 7-9 the minimum settling rate of 1,150 gpd/ft^2 was observed at a MLSS of 2,600 mg/l. To provide for peak flow, the clarifier should be designed for an overflow rate of 770 gpd/ft^2, or less. It should have a 12 ft effective water depth, and might be provided with a surface skimmer.

Maximum SVI at 2,600 mg/l MLSS = 82 ml/gm (Figure 7-9)

Approximate return sludge concentration at worst condition

$$= \frac{10^6}{SVI} = \frac{10^6}{82} = 12{,}200 \text{ mg/l}$$

Solids balance equation where R is sludge return rate

$$(Q_{max} + R)(MLSS) \leq (R)(\text{return sludge conc.})$$

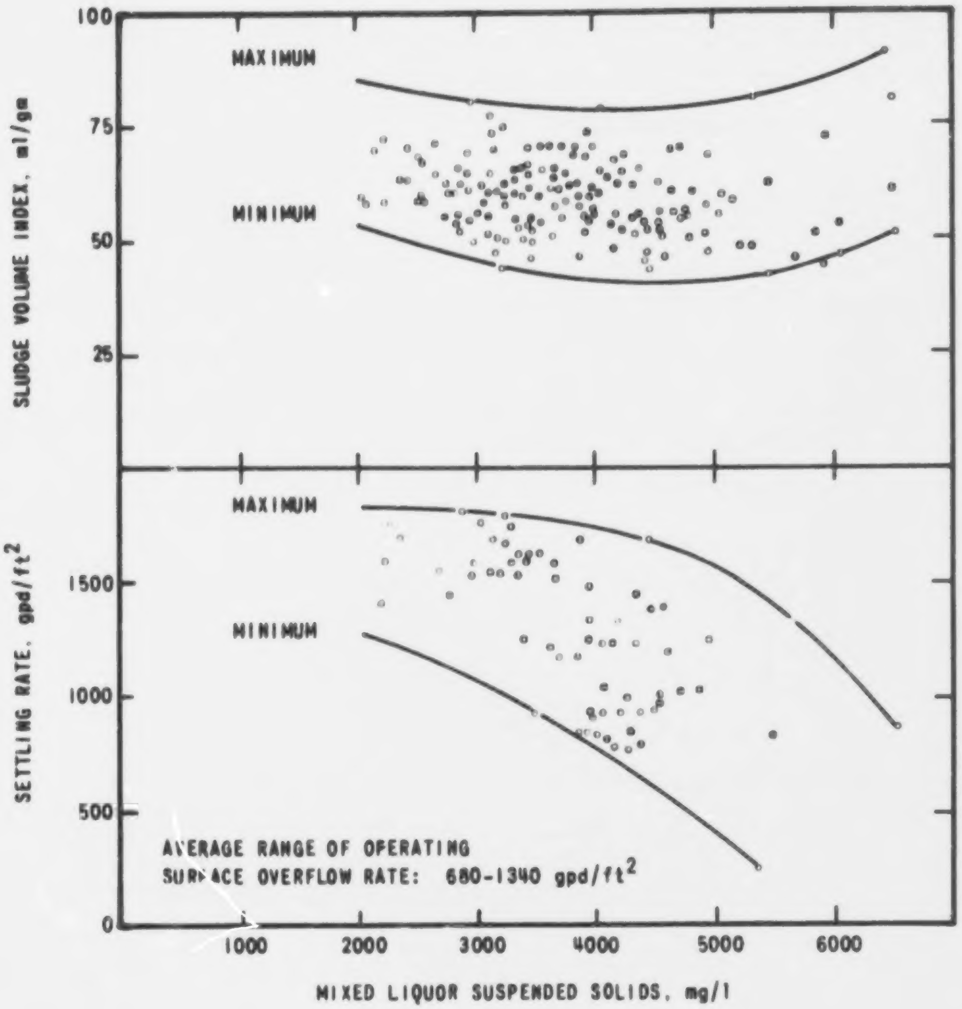

FIGURE 7-9 PHYSICAL CHARACTERISTICS OF HIGH RATE ACTIVATED SLUDGE WITH MINERAL ADDITION (8,16)

or considering just the case of equality,

$$R = 1.5\ Q_{ave}\ \frac{2{,}600}{12{,}200 - 2{,}600}$$

$$R = 0.4\ Q_{ave}$$

(Actually, effective design would indicate a ±50% capability; i.e., 0.2 to 0.6 Q_{ave} capability)

Solids to be wasted per day = 80 mg/l at a minimum sludge solids concentration = 12,200 mg/1. Assume continuous wastage with separate pump.

$$W = Q_{ave}\ \frac{80}{12{,}200} = \underline{0.0065\ Q_{ave}}$$

(Again, effective design would indicate a ±50% capability; i.e., 0.003 to 0.01 Q_{ave} capability).

5. **Alum Cost**

$$= (11.3)(0.32)(8.34)(10^{-3}) = \underline{3.1¢/1{,}000\ \text{gal}}.$$

Additional Comments

No mention has been made of the effect of sidestream phosphorus contributions. With conventional solids handling and treatment processes, gravity and flotation thickening, filtration and centrifugation, aerobic and anaerobic digestion, and incineration, phosphorus recycle should not be a problem and any that occurs will largely be associated with the solids fraction of the minor process streams. However, processes that cause biological solids liquefaction and hydrolysis, such as heat treatment processes, will result in a substantial recycle of soluble phosphorus with a corresponding increase of metal ion requirements.

Example No. 2: Same system as Example No. 1 but with sulfuric acid addition for pH adjustment (residual alkalinity = 30 mg/l as $CaCO_3$) and polymer addition for solids control.

Detailed calculations are not shown for the problem since the techniques are essentially the same as in Example No. 1. Results are summarized, along with results from the other examples, in Table 7-5. Highlights from this problem are as follows:

1. COD, BOD_5, and suspended phosphorus concentrations in the effluent are substantially reduced by an assumed polymer dose of 0.3 mg/1. The chemically removed phosphorus was calculated for this case by multiplying the 0.6 lb P removed/lb Al^{3+} added, from Figure 7-6, by 1.33. The 1.33 was obtained from Figure 7-4, assuming an adjusted pH of about 6. The result was a phosphorus residual of 0.1 mg/1 with an aluminum dose of 11.3 mg/1. But

Figure 7-8 showed from experimental data a nominal 0.5 mg/l residual phosphorus at this Al:P ratio and, therefore, the conservative residual of 0.5 mg/l soluble phosphorus was chosen.

Table 7-5

SUMMARY OF DESIGN CALCULATIONS

Item	Example 1	Example 2	Example 3	Example 4
pH Adjustment	No	Yes	Yes	Yes
Polymer Addition	No	Yes	Yes	Yes
Split Treatment*	No	No	No	Yes
Mixed Media Filtration	No	No	Yes	Yes
Total Aluminum Dose, mg/l	11.3	11.3	27.3	18.2
Effluent Quality, mg/l				
Suspended Solids	20	2	0	0
COD	50	39	38	–
BOD_5	15	11	11	–
Total Phosphorus	3.3	0.6	0.2	0.2
Total Soluble Phosphorus	2.3	0.5	0.2	0.2
Solids Production, mg/l				
Biological	57	57	57	57+
Chemical	43	45	92	66
Alkalinity Loss Due to Al^{3+} Addition mg/l as $CaCO_3$	30	21	109	59
SO_4^{2-} Increase Due to Al^{3+} Addition, mg/l	61	61	147	98
Alum Cost, ¢ / 1,000 gal.	3.1	3.1	7.4	4.9
Polymer Cost, ¢ / 1,000 gal.	0	0.3	0.3	0.3
Acid Cost, ¢ / 1,000 gal.	0	1.7	0	1.0
Total Cost, ¢ / 1,000 gal.	3.1	5.1	7.7	6.2
Clarifier Overflow Rate, gpd/ft^2	770	670	360	
Return Sludge Rate	0.4Q	0.6Q	0.9Q	
Waste Sludge Rate	0.0065Q	0.008Q	0.012Q	

*Split treatment here incorporates mineral addition as in Example No. 1 without pH adjustment, followed by mineral addition with pH adjustment at a later point.

2. The alkalinity loss due to hydrolysis was also reduced due to increased $AlPO_4$ formation. Using 96% strength sulfuric acid and assuming a residual alkalinity of 30 mg/l as $CaCO_3$, a calculated sulfuric acid requirement of 101 mg/l was found. The sequence of chemical addition is acid, alum, and polymer, with about 5 minutes of mixing at peak flows between each application.

3. Dissolved solids additions totaled 156 mg/l SO_4^{2-}: 61 mg/l due to the alum and 95 mg/l due to the acid.

4. Because of improved solids capture, the operating MLSS in the aerator increased to 3,200 mg/l necessitating a lower surface overflow rate in the clarifier, and greater return and waste sludge handling capability.

5. Chemical costs increased to 5.1¢/1,000 gal. with benefits in phosphorus, suspended solids, BOD_5 and COD removals.

Example 3: Assuming mixed media filtration provides complete solids control, what is the change in system calculations if a residual effluent phosphorus of 0.2 mg/l as P is to be provided?

Highlights from this problem are as follows:

1. Essentially no differences from Example No. 2 are found in effluent COD and BOD. However, the mixed media filter does provide positive effluent solids control in the event of system upset. An Al:P weight ratio of 3 (Figure 7-8) was used.

2. The addition of alum caused an alkalinity consumption of 109 mg/l, which would deplete the background alkalinity of the wastewater to 41 mg/l as $CaCO_3$. If a substantially higher alum dosage had been required, all of the alkalinity might have been consumed, and addition of lime might have been necessary.

3. Dissolved solids addition consisted of 147 mg/l SO_4^{2-} from the alum.

4. Additional solids productions resulted in an operating solids level of 4,800 mg/l in the aerator, necessitating a lower surface overflow rate and higher return and waste sludge capabilities than in earlier examples.

5. Total chemical treatment costs are raised to 7.7¢/1,000 gal.

Example No. 4: Split treatment (chemical addition at more than one point) is utilized for a required effluent phosphorus concentration of 0.2 mg/l. Assume the first point of addition is as in Example No. 1. Further mineral addition will be made to the effluent in Example No. 1 either before filtration or during a second stage of biological treatment.

1. Changes in effluent COD and BOD are not calculated since this is dependent upon whether split treatment is practiced with just filtration, or whether another biological system is added. In a real situation it would be better to reduce the aluminum dose below that in Example No. 1 in the first stage system for more efficient chemical use. This was not done here for design simplicity. Another possibility is addition of part of the chemical dose in the primary settler.

   The Al^{3+} dose of 6.9 mg/l for the second addition was arrived at using Figure 7-8 for a pH adjusted system.

2. The alkalinity depletion due to aluminum addition was calculated as 59 mg/l as $CaCO_3$. This would leave a residual alkalinity of 91 mg/l. Sixty mg/l as H_2SO_4 was added to reduce the alkalinity to 30 mg/l. Again, in a real situation acid might not have to be added to this wastewater.

3. Dissolved solids additions were calculated as 98 mg/l SO_4^{2-} due to alum addition and 59 mg/l SO_4^{2-} due to acid addition.

4. Clarifiers were not sized in this example. If two biological processes were used in series, the final clarifier of the first process could be designed with an overflow rate from the middle of the performance curve in Figure 7-9.

5. The cost of acid is 1.0¢ / 1,000 gal. Total chemical costs are 6.2¢ / 1,000 gal.

Comment: This final example, in comparison to example No. 3, shows the benefit derived from split chemical treatment for low phosphorus residuals.

7.5 References

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Chapter 8

PHOSPHORUS REMOVAL BY LIME TREATMENT OF SECONDARY EFFLUENT

8.1 Description of Process

8.1.1 Theory

Lime treatment of wastewater is essentially the same process as the familiar lime softening of drinking water supplies. The objectives, however, are quite different. While softening may occur, the primary objective is to remove phosphorus by precipitation as hydroxyapatite. This reaction was described in Chapter 3.

During phosphorus precipitation other important reactions occur. The reaction of lime with alkalinity, that results in calcium removal when carrying out softening, not only takes place when treating wastewater, but may have a very important effect on the general efficiency of the process. This reaction can be considered to take place in the two following ways:

$$Ca(HCO_3)_2 + Ca(OH)_2 \rightarrow 2\ \underline{CaCO_3} + 2\ H_2O$$

$$NaHCO_3 + Ca(OH)_2 \rightarrow \underline{CaCO_3} + NaOH + H_2O$$

The first equation is that for softening. Some wastewaters do not contain enough calcium, however, for that equation to be satisfied. Calcium carbonate precipitation may still occur, but by the second reaction. The reactions to form $CaCO_3$ are important for two reasons: the lime consumption determines to a considerable extent the lime dose required for operating the process, and the resulting $CaCO_3$ acts as a weighting agent to aid in settling of sludge.

Another reaction that may be important is precipitation of $Mg(OH)_2$ as follows:

$$Mg^{2+} + 2\ OH^- \rightarrow \underline{Mg(OH)_2}$$

This reaction does not approach completion until the pH is raised to 11. Magnesium hydroxide is a gelatinous precipitate which aids colloid removal, but which hinders sludge thickening and dewatering. Where the pH must be raised to 11 or above to meet a phosphorus removal requirement or some other treatment objective, $Mg(OH)_2$ formation must be considered.

In the two-stage lime treatment process which will be described below, it is necessary to include recarbonation after the first stage to reduce the pH and precipitate the excess lime as $CaCO_3$ in the second stage. The following reaction occurs:

$$Ca^{2+} + CO_2 + 2\ OH^- \rightarrow \underline{CaCO_3} + H_2O$$

Carbon dioxide may also be used to lower pH after lime treatment. The important reaction in this case would be the conversion of Co_3^{2-} to HCO_3^-.

A final reaction that is of major concern, where lime is to be recovered by sludge recalcination, is as follows:

$$CaCO_3 \xrightarrow{\Delta} CaO + CO_2$$

The CaO produced would then be slaked to form $Ca(OH)_2$ before use.

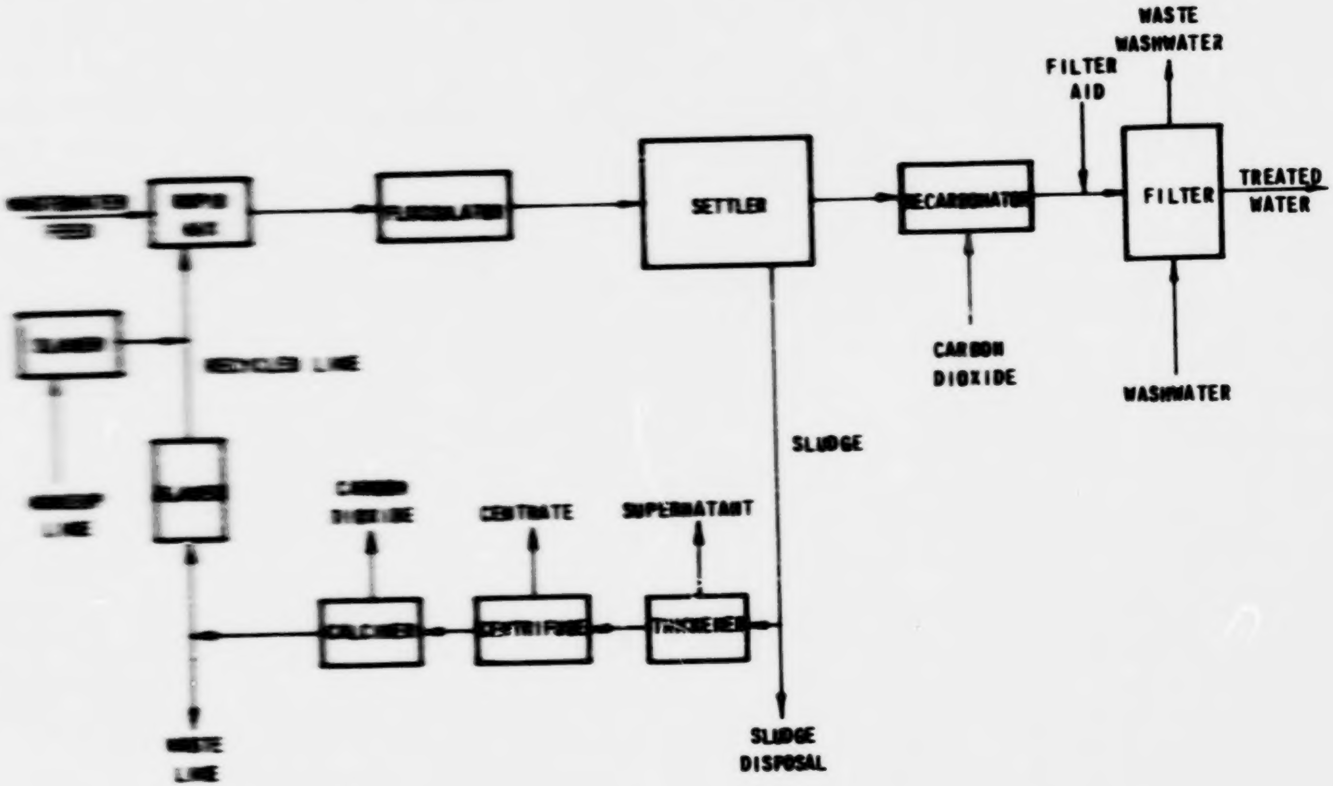
8.1.2 Treatment Systems

Two lime treatment systems may be used with wastewater, single-stage and two-stage. Figure 8-1 shows a single-stage system. In single-stage treatment lime is mixed with feed water to raise the pH to a desired value. Although the pH will depend upon the required phosphorus removal, it is likely to be substantially less than 11, and may be less than 10. Precipitation of phosphate and other materials, as indicated by reactions discussed above, takes place. Time is allowed in the appropriate equipment for precipitate particles to flocculate to sufficient size for good settling. The clarified water from the settler may be discharged directly or may be filtered to improve solids removal. Adjustment of pH with CO_2 may be necessary before discharge and will almost certainly be required before filtration to prevent post-precipitation of $CaCO_3$ from the unstable water. The settled lime sludge may be disposed of as landfill or may be recalcined for recovery of lime. In the latter case, the sludge is thickened, dewatered by centrifuge or vacuum filter, and calcined. The calcined product is then slaked and reused. To avoid buildup of inerts in the lime, some of the sludge or recalcined lime must be wasted or the inerts must be separated from the sludge before calcination.

Two-stage treatment is somewhat more complicated than single-stage treatment. In typical two-stage treatment, shown in Figure 8-2, enough lime is added to the water in the first stage to raise the pH above 11. Precipitation of hydroxyapatite, $CaCO_3$, and $Mg(OH)_2$ occurs. Consideration of the solubility product for $CaCO_3$ and the equilibrium between CO_3^{2-} and HCO_3^- shows that the minimum solubility for Ca^{2+} occurs at a pH of about 10. At a pH of 11 or above there is a considerable concentration of Ca^{2+} present in the water. In two-stage treatment CO_2 is added after the first-stage settler to bring the pH down to about 10 where $CaCO_3$ precipitation results. The $CaCO_3$ is settled out and the clarified water is either discharged or sent to filtration. As in the case of single-stage treatment, pH reduction may be necessary before discharge and probably would be required before filtration.

8.2 Typical Performance Data

A number of full-scale and pilot-scale tertiary lime treatment plants are in operation and more full-scale plants are beginning operation. Only one plant, that at South Lake Tahoe, California, has operated for a significant period with recalcination of sludge for lime recovery. Most aspects of this 7.5 mgd plant have been discussed in detail by Culp and Culp (1). Although the data on handling of wastewater sludges for recalcination and the recalcination process itself are limited mainly to this plant,


WASTEWATER FEED
RAPID MIX
FLOCCULATOR
SETTLER
RECARBONATOR
FILTER AID
WASTE WASHWATER
FILTER
TREATED WATER
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WASHWATER
SLAKER
RECYCLED LIME
MAKEUP LIME
SLAKER
SLUDGE
CARBON DIOXIDE
CENTRATE
SUPERNATANT
CALCINER
CENTRIFUGE
THICKENER
WASTE LIME
SLUDGE DISPOSAL

FIGURE 8-1 SINGLE STAGE LIME TREATMENT SYSTEM

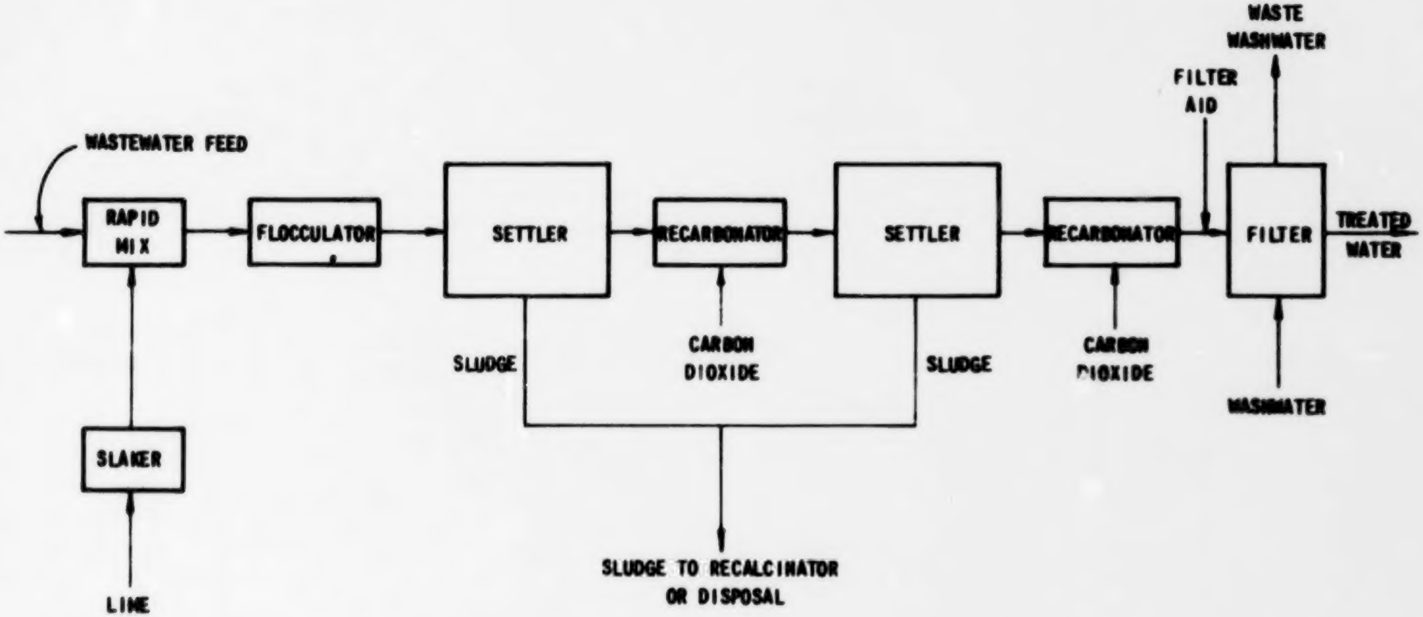
WASTEWATER FEED
RAPID MIX
FLOCCULATOR
SETTLER
RECARBONATOR
SETTLER
RECARBONATOR
FILTER AID
WASTE WASHWATER
FILTER
TREATED WATER
WASHWATER
CARBON DIOXIDE
CARBON DIOXIDE
SLUDGE
SLUDGE
SLAKER
LIME
SLUDGE TO RECALCINATOR OR DISPOSAL

FIGURE 8-2 TWO STAGE LIME TREATMENT SYSTEM

performance data on other parts of a lime treatment system are available from more than one location. Extensive data have been reported from pilot plants located in Washington, D. C., (2) and Lebanon, Ohio (3).

The lime treatment system at Tahoe is a two-stage clarification system which is followed by pressure filters of the multimedia type. These contain 3 ft. of a mixture of coal, sand, and garnet. The water passes through two filters in series. The first stage clarifier is operated at a pH of 11. To reach that pH requires about 300 mg/l CaO. A small amount of polymer is used to improve flocculation. The pH is reduced to 9.6 by recarbonation before the second-stage settler. Before filtration, the pH is further reduced to 7.5. A small amount of alum (1 to 20 mg/l) or a combination of alum and polymer is used as filter aid. Filter run length has varied between 4 and 60 hours.

Phosphorus removal at this plant always has been good and has improved as operating experience has increased. By returning plant waste streams containing precipitated phosphorus to the first stage flocculator, it is now possible to obtain routinely an effluent with less than 0.1 mg/l P. Before the filters the phosphorus concentration is about 0.4 mg/l.

In addition to phosphorus removal, there is significant removal of organic materials. During a period of intensive study of the first stage of clarification (unpublished data) 77% removal of BOD and 61% removal of COD were obtained. Although there was not a large removal of suspended solids, there was a significant change in the character of the solids from organic to largely inorganic. Further removal of organic materials and suspended solids occurs over the remainder of the system. Culp and Culp report typical filter effluents with a BOD of 3 mg/l, COD of 25 mg/l, and turbidity of 0.3 JTU.

Sludges from the first stage settler and the settler following recarbonation are sent to a gravity thickener. Solids concentration increases during thickening from about 1% to from 8 to 20%. Thickened sludge is then centrifuged to form a cake with from 30% to more than 40% solids. The cake is next calcined in a multiple hearth furnace and the recovered lime slaked for reuse. Initially it was planned to operate the centrifuge for high solids recovery in the cake. This necessitates discarding a significant amount of the recovered lime to prevent buildup of precipitated phosphate and other inerts. Since operation began, it has been found that much of the phosphate and some other inert materials can be separated from the $CaCO_3$ in the sludge by operating with a rather high solids concentration in the centrate. Approximately 90% of the phosphorus can be removed in the centrate, for example, when 25% of the solids entering the centrifuge are allowed to remain in that stream. This classification procedure results in a loss of about 15% of the recoverable lime. A greater degree of utilization of recalcined lime compensates for the loss, however, and the load on the calcining furnace is reduced. Over three years of operation, mostly without classification in the centrifuge, the average concentration of CaO in the recalcined product was 66%. Recalcined lime made up 72% of the total used at the plant.

Considerable operating data have been obtained by O'Farrell and Bishop (2) on a two-stage system operated at the EPAWQO-DC Pilot Plant in Washington, D.C. This 50,000 gpd plant treated secondary effluent from a pilot activated sludge system at the municipal treatment plant. The first stage pH was maintained at about 11.7, using a lime dose of about 400 mg/l as CaO, and the pH after recarbonation was maintained at about 10.3. Ferric iron was added at a rate of 5 mg/l in the second stage to improve flocculation. Water from the second-stage settler was filtered without further pH adjustment through gravity-flow dual-media filters consisting of anthracite coal over sand. The average filter run length was more than 50 hours.

Phosphorus removal for the system was similar to that obtained at Tahoe; 0.09 mg/l as P was the average concentration remaining. Phosphorus concentration before filtration averaged 0.13 mg/l.

In addition there was significant organic and suspended solids removal. The average BOD was reduced from 15 mg/l to 2.1 mg/l before and 1.5 mg/l after filtration. Suspended solids were reduced from 33 mg/l to 17 mg/l before filtration and 3.8 mg/l after filtration.

Sludge from the second stage settler was returned to the first stage and all sludge removed from the system was from the settler of the first stage. This sludge contained about 5% solids and constituted about 1.5% of the feed volume. Although there was sludge thickening and calcining equipment available, only preliminary study was made of lime reuse. Limited results showed improved thickening of the sludge when recalcined lime was recycled to the system.

A 75 gpm single-stage pilot system was operated at the Lebanon, Ohio Sewage Treatment Plant. Feed water was activated sludge effluent. This system, consisting of a flocculator-clarifier followed by dual-media filters of anthracite coal and sand, is described by Berg, et al. (3). Sludge was gravity thickened and discharged to sand drying beds. The system has been operated over a pH range from about 9 to 11 with most data taken at a pH of 9.5. Before filtration the pH was reduced to about 8.8 with sulfuric acid to prevent precipitation on the filter media. Filter run length averaged about 90 hours.

Effluent phosphorus concentrations for the system are shown in Figure 8-3. The effect of pH is clearly indicated. Since the average phosphorus concentration of the secondary effluent was about 10 mg/l as P, approximately 95% removal was obtained at a pH as low as 9.5. Before filtration the average phosphorus concentration was 0.75 mg/l at a pH of 9.5. Removal improved only slightly at higher pH because of the presence of precipitated phosphorus in the effluent.

Average suspended solids concentration of the secondary effluent was 43.5 mg/l. This was reduced to 16.5 mg/l in the clarifier. Much of the suspended matter in the clarifier effluent consisted of inorganic precipitates. After filtration the water was of high clarity. During a six month period when the biological treatment plant operated well, turbidity averaged 0.2 JTU. Even when the suspended solids load from the activated

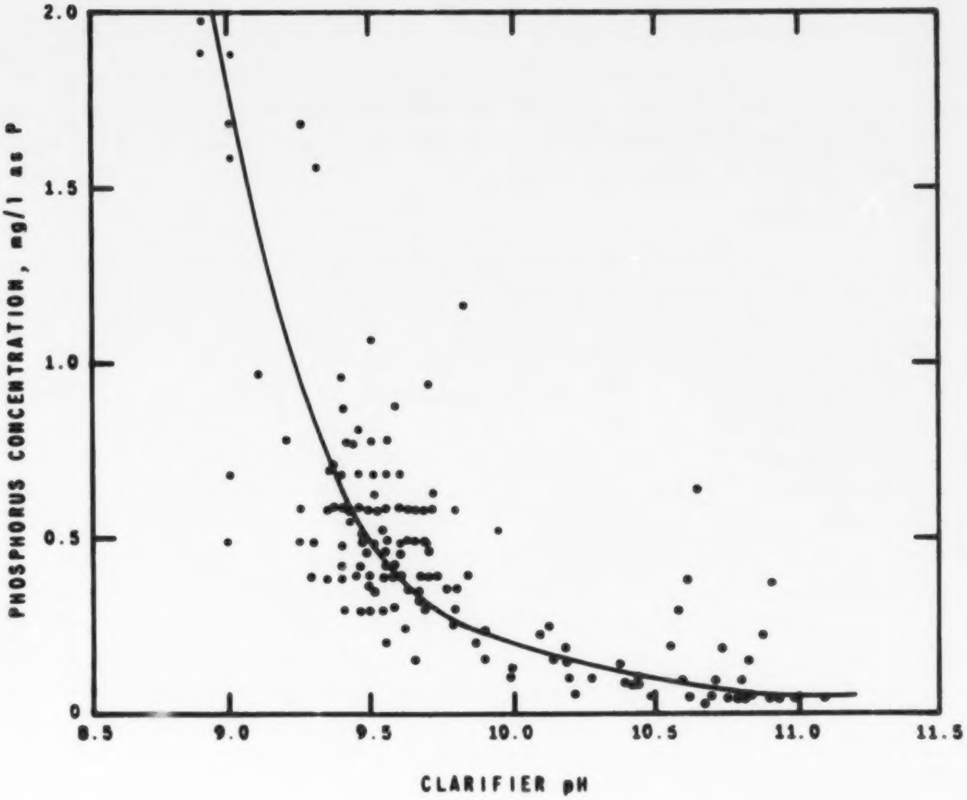

FIGURE 8-3 EFFECT OF pH ON PHOSPHORUS CONCENTRATION OF EFFLUENT FROM FILTERS FOLLOWING LIME CLARIFIER

sludge plant was high, the lime treatment system produced water with a daily average turbidity not exceeding 2.5 JTU.

During a two month period in which extensive measurement of organic removal was made, total organic carbon was reduced by 63% over the clarifier and 68% over the clarifier and filters. Organic carbon measurements taken at other times indicated removals of from 55 to 74%. Occasional BOD and COD samples indicated removals of 86% and 62%.

Sludge was usually removed from the settler at 1.5 gpm giving a sludge concentration of 2.8% solids. This was thickened to about 10% before being pumped to drying beds. The sludge dewatered quickly to about 50% solids. Equipment was not available for recalcining the sludge.

The lime requirement is an important consideration in lime treatment. This can vary over a wide range depending on operating pH and water composition. From reactions given earlier it was seen that alkalinity has an important effect on the lime dose. Buzzell and Sawyer (4) have shown, for example, that for several wastewaters the lime dose required to reach a pH of 11 correlated approximately with alkalinity. Examination of the earlier equations would show also that calcium hardness can affect lime dose. One part by weight of CaO can react with from 0.89 to 1.79 parts of bicarbonate alkalinity expressed as $CaCO_3$, the lower value applying to very soft waters and the higher value to very hard waters. In addition to the reaction of lime with hardness, other competing reactions occur in lime treatment of wastewater. Also, there may be incomplete reaction of the lime. All of these complications make calculation of lime dose difficult. The result is that, at present, determination of lime dose is largely empirical. Approximate values have already been given for the plants at Tahoe and Washington, D.C. Some approximate values for the Lebanon work are shown in Table 8-1. It appears from the data already available that the lime dose will usually be in the range of 300 to 400 mg/l as CaO for two-stage treatment, and from 150 to 200 mg/l where single-stage treatment is satisfactory.

Table 8-1

LIME REQUIREMENTS

Feed Water Alkalinity (mg/l as $CaCO_3$)	Clarifier pH	Approximate Lime Dose (mg/l of CaO)
300	9.5	185
300	10.5	270
400	9.5	230
400	10.5	380

8.3 Criteria for Selection of Process

Lime treatment of secondary effluent represents a significantly higher capital cost and somewhat higher operating cost for phosphorus removal than mineral addition to a conventional treatment plant. Lime treatment has, however, several advantages over mineral addition. It can be considered more dependable since it adds additional flocculation and sedimentation steps to the system. Upsets in the conventional plant, which would reduce the efficiency of phosphorus removal by mineral addition, would have less effect on tertiary lime treatment. Because tertiary lime treatment is separate from the conventional treatment plant, it adds flexibility to operation of the system. For very high degrees of phosphorus removal, lime treatment would be the method of choice, not only because of the inherent greater dependability mentioned above, but because of the ability to produce an effluent slightly lower in phosphorus content. Lime treatment decreases the total dissolved solids content of the water by removal of hardness and alkalinity while mineral addition adds to the total dissolved solids. In the case of alum addition, for example, 5.3 parts by weight of sulfate are added for each part of aluminum. Lime treatment has the capability of removing turbidity to very low levels. Where there are plans for recreational reuse or certain industrial reuses of the treated water, low turbidity along with low phosphorus content of the lime treatment effluent is very desirable. Lime treatment offers the opportunity for recovery of the treatment chemical. At the present time there are no acceptable methods for recovery of aluminum or iron salts.

The choice of single-stage or two-stage lime treatment depends partly upon the degree of phosphorus removal required, but more importantly, on the alkalinity of the water. Unless a high treatment pH is used, waters with low alkalinity, in the range of 150 mg/l as $CaCO_3$ or less, form a poorly settleable floc because of the low fraction of dense $CaCO_3$. A pH above 10 is needed even to obtain measurable $CaCO_3$ production. A pH of 11 or above is then needed to precipitate $Mg(OH)_2$ which aids in settling of fine particles. Since neither discharge nor reuse of the high-pH water is likely to be acceptable, addition of a second treatment stage after recarbonation to a pH of about 10 becomes generally necessary. It would be possible to recarbonate to a much lower pH and avoid the second treatment stage, but this would result in a high calcium effluent and would eliminate the chance to produce high $CaCO_3$ sludge in situations where lime recovery was contemplated. Although most of the phosphorus removal occurs in the high pH first stage, in accordance with the pH dependence of phosphorus solubility as shown earlier by Figure 8-3, some removal does also occur in the second stage. Another important reason for including the second stage is to assure better control of clarification. With low alkalinity waters there is sometimes difficulty in settling the sludge in the first stage even at high pH. The second stage settler with its high $CaCO_3$ sludge, prevents solids carryover when the first stage settler does not operate properly.

With high alkalinity waters, a well settling floc is formed at pH values as low as 9.5. There is no need for two-stage treatment unless the required degree of phosphorus removal necessitates a high pH. In addition to obtaining a small degree of phosphorus removal, the second stage would be used in these cases to lower calcium content in the effluent and to obtain high $CaCO_3$ sludge for lime recovery. There is not yet available operating experience at enough plants to state positively the alkalinity at which single-stage treatment will perform satisfactorily, from the standpoint of floc settleability. At present, experience indicates that at an alkalinity of 150 mg/l as $CaCO_3$, single-stage treatment probably cannot be used. Between 150 mg/l and 200 mg/l settleability will probably depend on the amount of organic floc present. Above 200 mg/l, settleability is likely to be satisfactory.

8.4 Description of and Criteria for Choice of Equipment

8.4.1 Single-Stage System and First Stage of a Two-Stage System

Diagrams for single and two-stage lime treatment systems are shown in Figures 8-1 and 8-2. The first part of each system is made up of clarification equipment essentially the same as that found in water treatment plants. It includes a chemical feeding system (see Chapter 10), a rapid mix tank, a flocculator, a settler, and a means for sludge removal. Mixing, flocculation, and settling may be carried out in separate vessels or tanks, or they may be combined in one integrated unit. The system consisting of separate tanks has been used for many years in water treatment and was the system adopted for the first stage of the Tahoe lime treatment plant. The integrated type of equipment is sometimes referred to as an upflow clarifier or a sludge blanket clarifier. The latter term can be misleading, however, since not all such units operate with a sludge blanket. A diagram of a unit that does not operate with a sludge blanket is shown in Figure 8-4. Culp and Culp (1) recommend the system of separate tanks because that system allows separate control of each part of the process. They point out that such a system has greater flexibility in points of addition for chemicals. They also point out that in the integrated units with a sludge blanket there may be difficulty in controlling the blanket height and the blanket may become anaerobic, leading to poor phosphorus and solids removal. Pilot studies at other locations have shown that excellent solids removal can be obtained with sludge blanket equipment, but that blanket instability could be a problem. On the other hand, pilot upflow units designed to operate without a sludge blanket have proved to be very effective at Washington, D.C. (2). Results available at this time indicate that both the system of separate tanks and the integrated units without sludge blankets should be considered for design of new plants.

The Tahoe plant is an excellent example of a system with the first stage consisting of separate tanks for each unit process. At the present time the plant must be relied upon heavily as a guide to design of such systems. Description of the equipment and some design parameters are given by Culp and Culp. The rapid mix tank which is located in one corner of the flocculator has about a 30 second residence time at design flow. (The plant usually runs with a daily average flow of about one third of design flow.) Mixing is accomplished with a vertical shaft mixer. The flocculator is a

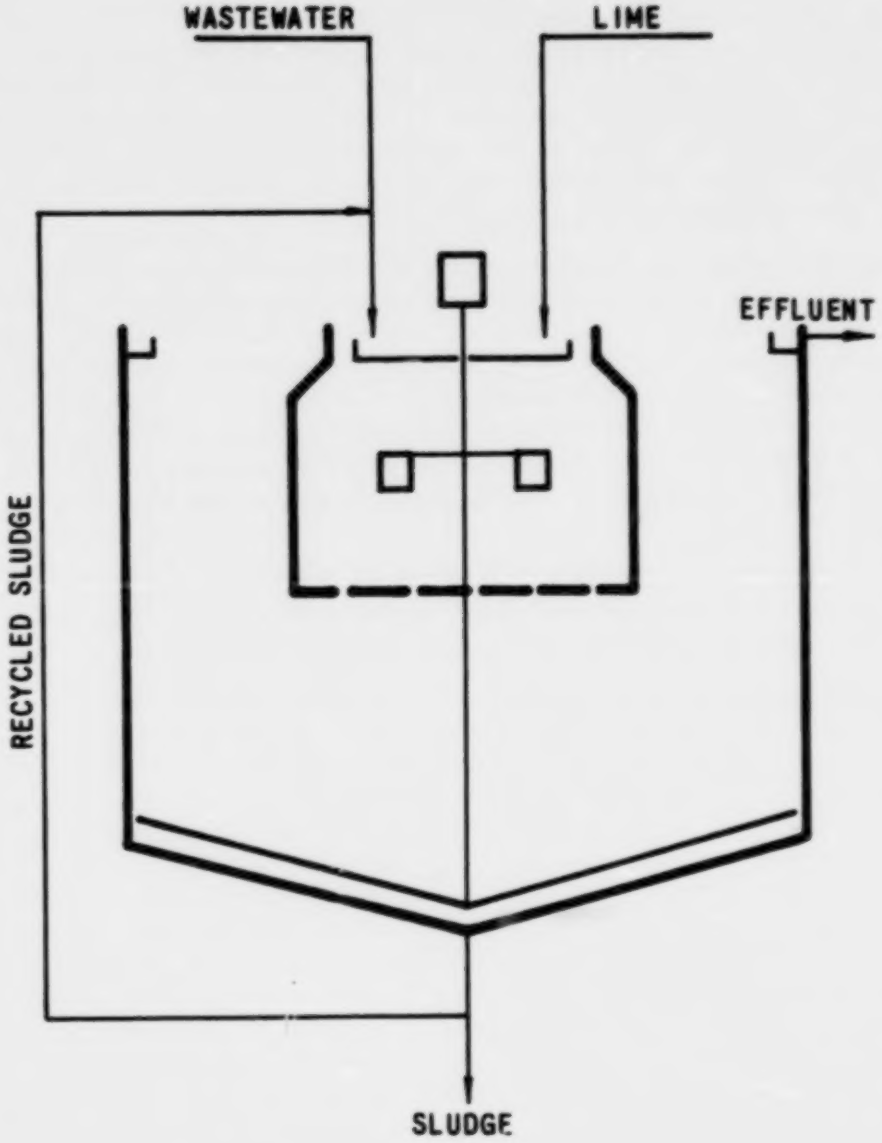
WASTEWATER
LIME
EFFLUENT
RECYCLED SLUDGE
SLUDGE

FIGURE 8-4 TYPICAL UPFLOW CLARIFIER

square tank with a depth of 8 ft. At design flow the residence time is 4.5 minutes. The flocculator is provided with air agitation, but operating experience has shown that this agitation is not necessary. The settler is circular with a center inlet. The depth is 10 ft and the design overflow rate is 950 gpd/ft^2. There are two sludge pumps, one a variable-speed centrifugal, and the other a positive displacement Moyno. Provision has been made to return part of the sludge to the rapid mix tank. The concept of returning sludge to the section of the clarification equipment where precipitation is taking place is called solids contact. The objective is to hasten precipitation and to obtain larger precipitate particles which will settle well.

The ease with which flocculation occurs at Tahoe would suggest that a flocculator for this service presents no design problems. Experience is, however, limited. Pilot testing would be very desirable to determine the flocculating characteristics of a particular wastewater. Jar tests may be of some value. Unfortunately, the effect of solids contact, which experience at Tahoe and elsewhere indicates is beneficial, is difficult to duplicate in a jar test. If jar tests indicated good flocculating characteristics without benefit of solids contact, however, flocculation should be as good or better with solids contact. When pilot studies are carried out, provision should be made for solids contact.

Comments concerning the effect of solids contact in jar tests also apply to settling rates determined from these tests. Rapid settling without benefit of solids contact would be a strong indication that settling with solids contact would also be rapid.

Clarification equipment of the integrated type for use in municipal water treatment is manufactured by a large number of companies. Units of this type used for treatment of wastewater have been essentially of the same design as used for municipal water supplies. The basins are usually circular with the rapid mix and flocculating sections located at the center of the settler. Agitation for mixing and flocculating as well as power for sludge scrapers is provided at the center of the settler. Solids contact is usually provided, and may be carried out with best control by external circulation of sludge to the mixing section. In some designs, however, internal recirculation is used. There is considerable variation in the geometry and complexity of the mixing and flocculating sections depending upon the manufacturer. The unit shown in Figure 8-4 is a relatively simply type. Units from three different manufacturers were used in pilot work at Lebanon, Ohio and Washington, D.C. These and others at additional locations have performed effectively. However, in some cases with operation of a sludge blanket, difficulties have been encountered. A sludge blanket can be eliminated by avoiding designs in which the wall separating the center compartments from the concentric settler reaches close to the bottom of the settler. The design in Figure 8-4 prevents sludge blanket formation.

There has been little effort to modify this type of equipment from use in water treatment to use with wastewater. Experience at Tahoe would suggest, for example, that less flocculation time, and probably less agitation, is required than is generally provided. A unit such as that shown by Figure 8-4 may have about 40 minutes flocculation time at design capacity for water treatment, nearly nine times that at

Tahoe. Whether modifications in such equipment would be worthwhile remains to be seen.

If a designer chooses to use equipment that is on the market, he does not have complete flexibility in specifying the design. Fortunately, equipment satisfactory for water supply treatment has proved satisfactory for wastewater, except that a lower settler overflow rate is necessary. The single-stage treatment system at Lebanon, Ohio was operated at a constant overflow rate of 1,440 gpd/ft^2. This proved satisfactory even with a sludge blanket. At Washington, D.C., tests were made with settler rates as high as 1,950 gpd/ft^2 although the average rate was 1,120 gpd/ft^2. For design, a peak, dry weather overflow rate of 1,200 to 1,400 gpd/ft^2, with the average somewhat lower, appears reasonable.

8.4.2 Recarbonation

In a two-stage system recarbonation follows the first stage settler. In a single-stage system and following the second stage of a two-stage system, pH reduction by recarbonation may be necessary to make the effluent suitable for filtering or for discharge. An excellent discussion of all aspects of wastewater recarbonation is given by Culp and Culp (1). The reader is referred to that publication, as well as one by Haney and Hamann (5), especially for information on sources of carbon dioxide. Sources include stack gas from sludge incinerators and lime recalciners, CO_2 generators, and commercial liquid CO_2.

The equipment used for contacting the CO_2 and water may be simply a tank with the gas being bubbled through the water. This is the system used at the Tahoe plant where the height of water over the CO_2 source is 8 ft. Pilot studies at Washington, D.C., with a tank depth of 11 ft and a turbine mixer to reduce bubble size and distribute bubbles, showed almost 100% absorption of CO_2.

When considering residence time, the recarbonation tank following the first stage of a two-stage system must be differentiated from the recarbonation tank used just to reduce effluent pH. The residence time of the recarbonator between the first and second stages is not particularly important. At Tahoe the recarbonation tank is only large enough to give a 5 minute residence time at design flow. In the Washington, D.C. studies, the residence time was 15 minutes. A problem that may arise when recarbonating wastewater is foam formation. This is most pronounced when the flow of bubbles is concentrated in parts of the recarbonation tank. The CO_2 distributing system should cover as well as possible the whole lateral area of the tank bottom. This would be especially true if the residence time were as low as 5 minutes.

In the recarbonation tank used just for effluent pH adjustment, sufficient residence time must be allowed for completion of reactions taking place. Culp and Culp recommend 15 minutes. In the Tahoe plant only 4 minutes is provided at design flow, but the water flows to storage ponds with a much longer residence time before the water is filtered. Just as in the case of recarbonation between stages of a two-stage system, good bubble distribution is important for prevention of excessive foaming.

Since one reason for using a two-stage system is to prevent a high calcium concentration in the effluent, the pH to which the water should be recarbonated between stages is that for maximum conversion of the calcium to $CaCO_3$. Where lime recovery is practiced, maximum formation of $CaCO_3$ is also desirable because it results in maximum CaO production. The optimum pH is about 10. At Tahoe, for example, 10.3 was selected because tests showed that pH to give maximum $CaCO_3$ production.

The pH to be selected when the objective of recarbonation is just pH reduction, will depend on a number of factors. An effluent standard may determine this pH. If stabilizing the water to prevent post precipitation is the objective, the pH may be determined by consideration of the Langelier Index. If, as in the case of Tahoe, a treatment step such as activated carbon adsorption is to follow lime treatment, this will determine pH. In the latter case a pH of about 7.5 has usually been selected.

The CO_2 dose requirements can be calculated from the chemical reactions taking place and a knowledge of the concentrations of the various forms of alkalinity in the water. For design of a new plant, alkalinity data must be obtained from samples of liquor taken from jar tests run at the same pH values as planned for the plant. Results must be considered very approximate, but they do help in sizing CO_2 feeding equipment. Culp and Culp discuss the calculations in detail. It was found in pilot plant work at Washington, D.C. that the CO_2 dose for recarbonation between the stages of a two-stage treatment system can be calculated reasonably well by considering the reduction in calcium concentration that occurs. This reduction is due to $CaCO_3$ formation with the CO_3^{2-} coming from the CO_2. The calcium content before recarbonation can be determined satisfactorily from jar tests run at the desired operating pH. The calcium content after recarbonation can also be determined approximately by jar test. Results from several pilot plants indicate the soluble calcium content at the pH of minimum solubility should be about 40 mg/l as Ca. This figure is probably just as good for calculations as a figure obtained from a jar test. The CO_2 dose in mg/l is then equal to:

$$(\text{Ca reduction in mg/l}) \left(\frac{44}{40}\right).$$

A safety factor of about 20% should be added to the calculated dose to compensate for inefficiency in absorption.

8.4.3 Second Stage Flocculation and Sedimentation

In two-stage treatment, recarbonation is followed by the second stage flocculation and settling. Experience with equipment for the second stage of treatment with wastewater is limited. Two very different systems have been tested. The simplest is the system at Tahoe. In the Tahoe plant the equipment used is just a longitudinal settler with a 30 minute detention time and a 2,400 gpd/ft^2 overflow rate at design flow. No flocculation equipment is provided. Culp and Culp (1) refer to the settler as a reaction and settling basin. It has been found at Tahoe that a significant part of the recarbonation reaction actually occurs in the settler. This is shown by the pH decrease of about 0.7 unit that occurs between the recarbonation tank and the settler outlet. The clarification that results at Tahoe is unusually good for the simplicity of the equipment.

Jar test and pilot plant work at Washington, D.C. indicated that flocculation after addition of a flocculating aid was required to obtain satisfactory clarification of that water. The pilot tests were run in upflow equipment of the integrated type discussed earlier. Without a flocculating aid, fine $CaCO_3$ precipitate escaped the settler, even at a settler overflow rate as low as 1,100 gpd/ft^2. Using 5 mg/l of Fe^{3+}, good operation was observed at overflow rates as high as 1,950 gpd/ft^2. Hydraulic limitations prevented testing at higher rates. Apparently, the good contact between solids and water during flocculation was beneficial in reducing $CaCO_3$ supersaturation in the effluent. Precipitation in the filters following the second stage did not occur, and no pH reduction was required.

In this work the sludge from the second stage was returned to the mixing section of the first stage to serve as a weighting agent. In waters of low alkalinity or expected high loads of biological solids, this procedure should be considered.

It is difficult to recommend the amount of flocculation to provide and the overflow rate to use for second-stage treatment. Jar tests would be of little value because of the complication of first raising the pH with lime and then lowering with CO_2. Pilot testing may be desirable. Such a pilot system would have to include the first stage equipment and recarbonation equipment, although these could be of crude design. In the absence of pilot tests to indicate otherwise, flocculation should be provided in the second-stage system. If filtration is included, a peak dry weather overflow rate of up to 2,000 gpd/ft^2 in the settler may be used. If filtration is not provided, a somewhat lower overflow rate should be chosen to assure reasonable effluent quality during periods when an upstream part of the system is not operating properly.

8.4.4 Filtration

The last step in a complete lime treatment system is filtration. Although there are a variety of filter types that could be used, essentially all the recent test work has been done with downflow filters of multimedia type. These are essentially the same in design as a rapid sand filter, except that the media are graded from coarse at the filter surface to fine at the filter outlet. This is accomplished by using media of different densities with the largest particles being composed of the least dense material. The result is a filter bed which has far more capacity for removing suspended solids than an ordinary sand filter, without impairing effluent clarity. Both gravity and pressure filters have been used. Each has certain advantages. Generally, however, pressure filters are more appropriate at smaller plants.

Two media combinations have been tested extensively. These are dual-media of coal and sand, and tri-media of coal, sand, and garnet. Filters used at Lebanon, Ohio following a single-stage system (3) and filters at Washington, D.C. following a two-stage system (2) were of the dual-media type. Both contained 18 in. of anthracite coal over 6 in. of sand. Average particle sizes of the media at Lebanon were 0.75 mm and 0.46 mm; at Washington, D.C., 0.9 mm and 0.45 mm. At 2 gpm/ft^2 the average run length at Lebanon was about 90 hours using 8 ft of water as the terminating pressure loss. At Washington, D.C., filter run length averaged about 50 hours for an average rate of 3.4 gpm/ft^2 and a terminating pressure loss of 7 ft of water. High clarity waters were obtained in each case. Backwash in both cases was carried out at 20 gpm/ft^2. Backwash time was 5 minutes at Lebanon and 10 minutes at Washington, D.C. In the latter case, a surface wash was included.

These run length figures give a rough idea of the results that may be expected at other locations. For maximum run length at a given product quality, the depths of media and their particle sizes should be optimized. Although this was not done for the above-mentioned filters, pressure loss distribution in the filters at Lebanon as reported by Berg, et al. (3) indicates that good solids storage in the anthracite was being obtained. Small pilot filters could be used for optimization of the media.

A reasonable design rate for gravity-flow dual-media filters appears to be about 3 gpm/ft^2. Higher rates could result in inconveniently short runs during periods of high solids load. Rates much lower than 3 gpm/ft^2 result in excessive filter costs, not justified by the longer filter runs. Very long filter runs could result in backwash problems from biological activity in the filters. Recommendation of a 3 gpm/ft^2 rate is based upon having good operation of the second-stage settler. Frequent settler upsets must be avoided.

Tri-media filters are used at the Tahoe plant. These are pressure filters which at design flow operate at 5 gpm/ft^2. Each bed holds 3 ft of mixed coal, sand, and garnet as supplied by Neptune Microfloc. Two filters are used in series, and are backwashed in series, usually when the head loss reaches 16 ft of water. Run length has varied from 4 hours during heavy solids load to about 60 hours under good conditions. Backwash is carried out at 15 gpm/ft^2. The backwash water is reprocessed through the lime treatment system. More details are given by Culp and Culp (1).

From the available data, it is difficult to give precise criteria for choosing between dual-media or tri-media filters following lime treatment. The use of garnet allows for a smaller particle size at the bottom of the filter than is possible with sand. In case of floc weakness, the tri-media filter offers, therefore, more protection from turbidity breakthrough. The cost, however, is slightly higher. Where the tri-media fill is not used, filters can still be backwashed at the onset of turbidity breakthrough. Experience with dual-media filters on lime treated water has not shown sudden breakthrough to be an important problem. Where water of the highest clarity is required, tri-media filters may be of most value.

8.4.5 Sludge Handling

The sludge from a lime treatment system may be handled in two general ways. It may be thickened, dewatered, and disposed of or it may be thickened, dewatered, and recalcined to recover lime for reuse. For small plants, recalcination will not be economically competitive with disposal and should not be considered unless there are restrictions on disposal which make that alternative difficult. It can be assumed generally that for plants over 10 mgd, recalcination of lime sludge will be practical. Recalcination may also be practical at plants somewhat smaller than 10 mgd, depending on the cost of purchased lime and other local conditions.

Data given earlier indicate that the volume of sludge from lime treatment will vary from 1.5% to several percent of feed volume. Sludge concentration will probably be in the range of 1% to 5%. The actual weight of sludge will vary with the chemical composition and amount of suspended solids in the feed water. Values have been observed in the range of 4 to 7 lb/1,000 gal. Sludge production from the first stage of a two-stage system or from a single-stage system can be estimated approximately by weighing the dried sludge from jar tests run at the planned operating pH. Additional sludge produced in the second stage can be assumed to be the $CaCO_3$ formed from calcium concentration reduction during recarbonation. Calcium content before recarbonation can be obtained from the above-mentioned jar tests and after recarbonation can be assumed to be 40 mg/l as Ca.

Design information for the Tahoe sludge handling system is reported by Culp and Culp. (1) The reader is referred to that publication for further information. The Tahoe plant is the only one which includes recalcination of sludge from the lime treatment of wastewater, and that has been operated for a long period of time. Paramenters from that system can be used as a rough guide for design. The gravity thickener has a bottom scraper mechanism and is 8 ft deep. Overflow from the thickener is returned to either the primary clarifier or the first stage of lime treatment. The design solids loading is 200 lb/day/ft^2 and the design overflow rate is 1,000 gpd/ft^2. Some preliminary data from pilot work at Washington, D.C. suggest these loadings are high. The Tahoe equipment was sized, however, to handle a volume of sludge equal to about 9% of the plant design flow. The actual volume of sludge should usually be less than one third of that volume. At Washington, D.C., recalcined lime was found to produce a sludge which thickened significantly faster than sludge from virgin lime.

If lime is not to be recovered from the sludge, the underflow from the thickener can be placed directly on drying beds for final dewatering, or it can be dewatered by centrifuge or vacuum filter.

If sludge recalcination is planned, the thickened sludge would be dewatered, by centrifuging or filtering and fed to the calcining furnace. The centrifuge may have advantages over filters for this purpose, such as the ability to separate phosphate sludge from $CaCO_3$. Lime from the furnace is stored for reuse. It, along with any new lime, must be slaked before use. Sludge can be pumped from the thickener to the centrifuge. Cake from the centrifuge must be transported by conveyor. At Tahoe the centrate is sent to the primary clarifiers. It has been found at Tahoe that, by operating the centrifuge with less than maximum solids capture, much of the precipitated phosphorus can be retained in the centrate. This results in a higher quality lime. This mode of operation requires a second centrifuge to remove the phosphorus rich solids from the centrate. These solids may have value for use in fertilizer.

The Tahoe dewatering and recalcining system is described by Culp and Culp. The centrifuge is a 24 by 60 in. concurrent flow type. The calciner is a 14 ft-3 in. diameter, 6 hearth furnace operated at a top temperature of 1,850° F. Other types of furnaces have been used in water treatment plants and these should also be applicable to the sludge from wastewater treatment. The reader is referred to equipment manufacturers for further information about centrifuges and furnaces.

8.4.6 Control of Lime and Carbon Dioxide Feed

The feeding of lime to a lime softening system can be controlled in a number of ways. For the wastewater application, the most appropriate appears to be control of pH, with the pH value being selected for good suspended solids removal or to meet a phosphorus removal requirement. Where flow equalization is employed, pH sensing alone should be sufficient. Where there is substantial diurnal variation in flow, better control can be maintained by flow proportional-pH control. Systems are available for this type of control.

Carbon dioxide dose is also controlled by pH. As in the case of lime feed, flow proportional-pH control would be preferred where there is diurnal variation in flow.

There is a tendency for pH electrodes to become coated with precipitates and lose sensitivity. The precipitate can be dissolved with acid. In some instances, however, it has been found satisfactory to use manual control because of difficulties with the pH control system.

8.4.7 Scale Formation on Equipment

Because the water in a lime treatment system is supersaturated to some degree with $CaCO_3$ and other precipitating substances, there is a tendency for scale to form on equipment and pipe surfaces. The problem is particularly serious with the lime slurry from the slaker. This can quickly plug the slurry line to the rapid mix tank. There should be easy access to all parts of this line for cleaning. At one small plant a flexible hose was used to feed lime slurry. By periodically flattening or flexing the hose, the scale was removed. The mechanical mixer in the rapid mix tank will also become scaled and must be cleaned.

Scale can also form in sludge lines and the effluent line from the first-stage clarifier. It is recommended that open troughs be used wherever possible. Provision should be made for cleaning lines when open conduits are not possible.

Possible scaling of filters has already been mentioned. Recarbonation before the filters will minimize scale formation.

8.5 Capital and Operating Costs

Because of the short history of lime treatment of wastewater, there is a scarcity of capital and operating cost information. Costs are available for the Tahoe plant and are reported by Culp and Culp (1). Capital cost for the 7.5 mgd lime treating facility was $1,115,000 and cost of the filters was an additional $705,000. For 1969 the estimated operating costs exclusive of equipment amortization were 7.3¢/1,000 gal. for lime treatment without filtration and 2.8¢/1,000 gal. for filtration. Amortization of equipment, based on costs adjusted to the 1969 national average, interest of 5% for 25 years, and the assumption of the plant operating at full capacity, would add 2.7¢/1,000 gal. for treatment without filtration and 1.8¢/1,000 gal. for filtration. For the plant operating at full capacity the total cost of operation would then be 10.0¢/1,000 gal. without filtration and 14.6¢/1,000 gal. including filtration. The fraction of the cost resulting from amortization is a substantial 31% even with the assumption of full capacity. There is an obvious need to keep equipment size to a minimum. Flow equalization deserves strong consideration when planning for lime treatment of secondary effluent to minimize the initial cost of new equipment.

Additional costs for lime treatment based on information from Tahoe and other sources have been reported by Smith and McMichael (6). Tables 8-2 and 8-3 show costs based largely on their results. Capital costs have been updated to December, 1970. Amortization is at 6% for 25 years. For 1 mgd plants, recalcination equipment is

Table 8-2

CAPITAL COST OF LIME TREATING FACILITIES

Treatment		Cost ($) Plant Size (mgd)	
	1.0	10	100
Single-Stage without Filtration	100,000	1,200,000	5,500,000
Two-Stage without Filtration	160,000	1,500,000	7,900,000
Dual-Media Filtration	110,000	510,000	2,300,000

Table 8-3

TOTAL COST FOR LIME TREATMENT OF WASTEWATER

Treatment		Cost (¢/1,000 gal.) Plant Size (mgd)	
	1.0	10	100
Single-Stage without Filtration	13	7	4
Two-Stage without Filtration	16	9	6
Dual-Media Filtration	8	3	1.4

not included. All lime would be purchased. The applicability of recalcination was discussed earlier. Component costs making up the total cost figures represent an average for the whole country. Local conditions may cause significant deviation from these values.

8.6 References

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2. O'Farrell, T. P., and Bishop, D. F., "Lime Precipitation in Raw, Primary, and Secondary Wastewater", Presented at the 68th National Meeting of AIChE, Houston, Texas (March 1971).

3. Berg, E. L., Brunner, C. A., and Williams, R. T., "Single-Stage Lime Clarification of Secondary Effluent", *Wat. and Wastes Eng.*, 7:3, p 42 (1970).

4. Buzzell, J. C., and Sawyer, C. N., "Removal of Algal Nutrients from Raw Wastewater with Lime", *JWPCF*, 39:10, Part 2, p R16 (1967).

5. Haney, P. D., and Hamann, C. L., "Recarbonation and Liquid Carbon Dioxide", *JAWWA*, 61:10, p 512 (1969).

6. Smith, R., and McMichael, W. F., "Cost and Performance Estimates for Tertiary Wastewater Treating Processes", Robert A. Taft Water Research Center, Report No. TWRC-9 (June, 1969).

Chapter 9

PHOSPHORUS REMOVAL BY MINERAL ADDITION TO SECONDARY EFFLUENT

9.1 Description of Process

Alum and to a lesser extent iron salts have been evaluated as precipitants for phosphorus in effluents from conventional treatment plants. Both batch studies and continuous-flow systems up to 2.5 mgd have successfully reduced phosphorus to very low levels.

Equipment typically consists of mixers, flocculators, settlers, and filters. Some work has been done, however, with the chemicals added directly to filters. Specific studies are discussed later and provide additional information on equipment commonly used.

9.2 Summary of Design Information

An alum dosage of about 200 mg/l is required for phosphate removal from typical municipal wastewater while dosages of 50 to 100 mg/l are sufficient for effluent clarification. Iron salts, while successful in precipitating phosphate, have found little application because of residual iron remaining in the treated water. An Al:P molar ratio of 1:1 to 2:1 is required. The optimum pH for alum treatment is near 6.0 while for iron the optimum pH is near 5.0 (1). The pH of high alkalinity waters may be reduced either by using high dosages of alum or by adding supplementary dosages of sulfuric acid. Anionic polyelectrolytes are useful with alum in improving settleability of floc. Settling alone will reduce residual P to about 1 mg/l.

If higher removals are required, filtration must be employed. With proper operation, residual P may be reduced to less than 0.1 mg/l. Multimedia filters are preferred over sand filters because of the extended length of run. Microstrainers have not been successful in removing alum floc.

Surface overflow rates for clarifiers have ranged from 580 to 1,440 gpd/ft^2, but for consistent removal of phosphate, the lower end of the range is preferred. Filtration rates of 2 to 5 gpm/ft^2 are used.

No process has yet been developed for successfully recovering alum from sludge for reuse although both alkaline and acid regeneration were investigated (2, 3). The sludge can be dewatered by conventional methods. Because dewatering may be difficult, design of such facilities should be conservative.

Costs for tertiary treatment for phosphate removal range from 6 to 15¢/1,000 gal. depending on plant size and degree of treatment required.

9.3 Laboratory and Pilot Studies

9.3.1 South Lake Tahoe (3, 4, 5, 6)

Liquid alum was used to precipitate phosphate in a 2.5 mgd tertiary plant at South Lake Tahoe, California. The treated wastewater was applied to multimedia filters for solids removal without prior sedimentation.

Chemical feed was proportioned automatically to flow, and flash mixing was provided by an in-line mixer in the filter influent line. The filter media consisted of layers of anthracite coal, graphite, fine garnet and coarse garnet overlaying a supporting bed of gravel. Two filters were operated in series at 5 gpm/ft^2. Backwash with secondary plant effluent was initiated either by headloss or by a rise in turbidity of the effluent. Polyelectrolyte was used as a filter aid to control the depth of penetration of floc into the filter beds. With an alum dosage of 200 mg/l, 8 to 9 mg/l P in the secondary plant effluent was reduced to 0.03 to 0.3 mg/l P in the filter effluent. Turbidities were correspondingly reduced from 30 to 70 JTU to 0.2 to 3.0 JTU.

Efforts to regenerate alum from the sludge by both acid and alkaline treatment proved to be uneconomical and a lime treatment process was developed as a substitute for the alum process. Lime could be economically recovered for reuse and had the added advantage of producing a high pH which facilitated removal of nitrogen by ammonia stripping.

Total costs for phosphate removal only, as estimated in 1966, are shown in Table 9-1.

Table 9-1

1966 ESTIMATED COSTS FOR TERTIARY PHOSPHATE REMOVAL WITH ALUM

Plant Capacity (mgd)	Total Cost (¢/1,000 gal.)
2.5	11.7
10	8.5
50	7.1
100	6.6
200	6.3

9.3.2 Nassau County (7, 8)

Nassau County, New York, is evaluating treatment methods for its wastewater to provide a product water which is suitable for ground injection as a barrier to salt water intrusion. A pilot plant has been operated at 400 gpm producing a water which in most respects exceeds USPHS drinking water standards. Effluent from high rate activated sludge is pumped into a 40 ft diameter by 14 ft deep clarifier where alum and polyelectrolyte are added. Sludge is recirculated into the coagulation zone to improve efficiency of floc formation. The treated wastewater then passes downward through a flocculation zone and upward through a clarification zone to orifice type collectors. The clarified wastewater passes to 2 dual-media filters in parallel operated at 3.0 gpm/ft^2. Each filter consists of a 36 in. bed of No. 1-1/2 anthracite (effective size of 0.90 mm) above a 12 in. layer of sand (effective size of 0.40 mm). Filter backwash includes air scour, surface wash and high and low-rate backwashing. The filter effluent passes through 4 granular carbon contactors operating in series to remove organics.

Typical operating results with about 200 mg/l alum are shown in Table 9-2.

Table 9-2

NASSAU COUNTY PERFORMANCE DATA

	Influent	Effluent
Turbidity, JTU	20 - 25	0.2 - 1.5
Color, Pt-Co Units	20 - 40	none
COD, mg/l	78	13
P, mg/l	9	0.1 - 1.0

Costs have been estimated, for separate tertiary treatment only, at plant sizes of 1, 10, and 100 mgd as shown in Table 9-3.

Table 9-3

NASSAU COUNTY COST ESTIMATES

	Plant Capacity (mgd) (1)	(10)	(100)
	(Costs, ¢/1,000 gal)		
Process			
Coagulation	4.9	3.5	3.2
Filtration	1.8	1.1	1.0
Carbon adsorption	6.3	4.5	4.0
	13.0	9.1	8.2
Operating labor	28.0	5.6	1.8
Total Cost	41.0	14.7	10.0

9.3.3 Chicago, Hanover Park (9)

Alum was used to coagulate solids and precipitate phosphate from secondary effluent at Chicago's Hanover Park tertiary treatment plant. Treatment consisted of coagulation-sedimentation in circular clarifiers followed by either rapid sand filtration or microstraining for polishing. Average annual flow was 1.5 mgd.

Combined coagulation plus sand filtration removed 34, 58, and 82% P at dosages of 38, 60, and 140 mg/l of alum, respectively. Combined coagulation plus microstraining effected a lower phosphate removal of 23% at an alum dosage of 38 mg/l. In general, alum coagulation under the test conditions gave little or no improvement in suspended solids removal over that obtained by sand filtration of uncoagulated secondary effluent. The microstrainer did not significantly remove alum-coagulated solids.

9.3.4 Dallas (10)

Dallas, Texas, is operating a demonstration project designed to provide necessary design data for upgrading of the Dallas-White Rock trickling filter plant. Phosphorus removal facilities include a solids contact upflow clarifier unit and two multimedia filters. The clarifier has a rise rate of 0.5 gpm/ft^2 at the design flow of 100 gpm and a detention time of 4 hours. Both filters contain anthracite over sand media. One is a standard gravity filter. The other is a biflow type, with feed water entering at both the top and bottom of the bed. Filtered water is removed by a manifold-lateral collector located at the mid-depth of the bed. The gravity filter is designed for 50 gpm at a filtration rate of 4 gpm/ft^2. The biflow filter is designed for 100 gpm at a filtration rate of 8 gpm/ft^2, or 4 gpm/ft^2 for both the bottom and top of the bed.

The pilot plant process has not yet been optimized for phosphorus removal, however, limited data from two short test periods on trickling filter and activated sludge effluents are shown in Table 9-4.

Table 9-4

REMOVAL OF TOTAL ORTHOPHOSPHATE BY COAGULATION-CLARIFICATION IN THE DALLAS DEMONSTRATION PLANT

	Trickling Filter Effluent	Activated Sludge Effluent
Lime Dose, mg/l CaO	128	133
$FeCl_3$ Dose, mg/l Fe^{3+}	25	56
Soluble Phosphate In, mg/l P	6.5	7.0
Soluble Phosphate Out, mg/l P	0.43	0.21
% Soluble P Removed	93.5	97.0
Test Period, days	12	2

9.3.5 Dayton (11)

Dayton, Ohio, used alum in a 310 gpm pilot plant to remove phosphate and suspended solids from trickling filter effluent. The rectangular sedimentation basin was designed for 105 minutes detention and an overflow rate of 1,014 gpd/ft^2. Two filters designed for loadings up to 8 gpm/ft^2 were used for polishing effluent. One contained 30 in. of sand while the other used 20 in. of anthracite overlaying 10 in. of sand. The sand had an effective size of 0.40 to 0.45 mm with a uniformity coefficient of 1.35 to 1.70. The anthracite had an effective size of 0.85 to 1.20 mm with a uniformity coefficient of 1.35 to 1.80. Both air and water were used in backwashing.

Alum with activated silica or anionic polyelectrolyte as coagulant aids achieved phosphate removals at a dosage rate of 5 lb alum/lb PO_4^{3-} down to residuals of 4 mg/l PO_4^{3-} in filter effluents (89% removal of phosphate). Further removals down to 1 to 2 mg/l PO_4^{3-} required in excess of 200 mg/l alum. The clarifier was operated at overflow rates of 655 to 1,035 gpd/ft^2.

The filters removed 30 to 70% of the applied phosphate load. The dual media filter produced an effluent comparable to that from the sand filter with filter runs 2 to 3 times the length of the sand filter runs.

Costs to provide 90% phosphate removal with alum for the present 50 mgd flow in a 75 mgd plant were estimated at 6.5 to 8.2¢/1,000 gal.

9.3.6 Oxidation Pond Effluents

At Lancaster, California, a research project was initiated to renovate wastewater for use in recreational lakes(12). Existing treatment consisted of primary sedimentation followed by 45 to 60 days detention in oxidation ponds. For coagulation, laboratory jar tests showed alum to be more effective on a cost basis than either lime or ferric sulfate. Phosphate residuals were reduced to 1.1, 0.2, and 0.01 mg/l P with alum dosages of 150, 200, and 300 mg/l, respectively, at an optimum pH near 6.0. Pilot studies showed that both sedimentation and air flotation would provide satisfactory separation of most of the precipitated phosphates, however, filtration was necessary to obtain the required low residuals of less than 0.15 mg/l P.

Final specifications for a full scale process included coagulation with about 300 mg/l alum and 20 minutes of flocculation with a conventional paddle flocculator, sedimentation in a horizontal clarifier with an overflow rate of 400 gpd/ft^2, and filtration through a dual media gravity filter consisting of 18 in. of 0.5 mm anthracite over 8 in. of No. 20 sand.

Capital costs of a 0.5 mgd facility were estimated at $150,000 with operating costs of $184/$10^6$gal. The total cost of capital, operation, and maintenance for a 3 mgd facility was estimated at $150/$10^6$ gal.

Alum was recently evaluated by Shindala and Stewart(13) for phosphorus removal and polishing of waste stabilization pond effluents in Mississippi. Jar tests were used in the evaluation. Using a minimum of 90% removal of phosphorus and 70% removal of COD as criteria, the optimum alum dosage was 85 mg/l at pH 5.5. With higher doses, up to 97% phosphorus removal and 85% COD removal were observed. Ferric chloride and ferric sulfate were also tried but left a residual iron color in the supernatant.

9.3.7 Moving Bed Filter (14)

Alum and polyelectrolyte were used to treat trickling filter effluent in a proprietary moving bed filter at Liberty Corner, New Jersey. The device, a product of Johns-Manville Products Corp., used a bed of sand which is periodically moved upward in an inclined filter vessel. Clean sand is fed in at the bottom of the filter and floc-laden sand is removed mechanically at the top as required by pressure drop.

leaving a fresh filtration surface. The wastewater being filtered passes down through the bed and flows through screens at the sides of the filter body.

A major factor controlling phosphate removal is the molar ratio of Al:P. At an alum dosage of 200 mg/l and influent total phosphorus concentrations on the order of 25 to 28 mg/l as P (Al:P ratio = 0.6 to 0.7), the total phosphorus removal efficiency averaged 90%. With lower total phosphorus concentrations (Al:P ratios = 1.2 to 2.6) removal efficiencies averaged 95% and reached as high as 99%. The alum treatment also removed about 70% of suspended solids leaving an average of 15 mg/l in the effluent.

Capital costs for a 1 mgd moving bed filter are estimated at $264,000. Total operating costs for such an installation using 200 mg/l alum are estimated at 12¢/1,000 gal. treated.

9.4 References

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CHAPTER 10

STORAGE AND FEEDING OF CHEMICALS

10.1 Aluminum Compounds

The principal aluminum compounds that are commercially available and suitable for phosphorus precipitation are alum and sodium aluminate (1). Both are available in either liquid or dry forms. Alum is acidic in nature while sodium aluminate is alkaline and this may be an important factor in choosing between them. Aluminum chloride should be considered, and is discussed herein.

10.1.1 Dry Alum

10.1.1.1 Properties and Availability

The commercial dry alum most used in wastewater treatment is known as "filter alum" and has the approximate chemical formula $Al_2(SO_4)_3 \quad 14H_2O$ and a molecular weight of 594. Alum is white to cream in color. The pH varies between 3.0 and 3.5 in aqueous alum solutions having concentrations between 1% and 10%. Commercially available grades and their corresponding bulk densities and angles of repose are given in Table 10-1:

TABLE 10-1

AVAILABLE GRADES OF DRY ALUM

Grade	Bulk Density lb/cu ft	Angle of Repose
Lump	62 to 68	varies
Ground	60 to 71	43°
Rice	57 to 61	38°
Powdered	38 to 45	65°

Each grade has a minimum aluminum content of 17%, expressed as Al_2O_3. This corresponds to a 9% concentration as aluminum. Viscosity and solution crystallization temperatures are included in the subsequent section on liquid alum.

Since dry alum is only partially hydrated, it is slightly hygroscopic. However, it is relatively stable when stored under the extremes of temperature and humidity encountered in the contiguous United States.

The solubility of commercial dry alum at various temperatures is listed in Table 10-2:

TABLE 10-2

SOLUBILITY OF ALUM AT VARIOUS TEMPERATURES

Temperature °F	Solubility lb/gal
32	6.03
50	6.56
68	7.28
86	8.45
104	10.16

Dry alum is not corrosive unless it absorbs moisture from the air, such as during prolonged exposure to humid atmospheres. Therefore, precautions should be taken to ensure storage space is free of moisture.

Alum is shipped in 100 lb bags, drums, or in bulk (minimum of 40,000 lb) by truck or rail. Bag shipments may be ordered on wood pallets. In 1973, the price range for dry alum in bulk quantities is $60 to $65 per ton, F.O.B. the point of manufacture. Freight costs must be added to this. Bagging adds about $5 per ton to the base cost for bulk alum. Names and locations of major producers are in a subsequent section.

10.1.1.2 General Design Considerations

Water utilities usually use ground and rice alum because of their superior flow characteristics. These grades have less tendency to lump or arch in storage and therefore provide more consistent feeding qualities. Hopper agitation is seldom required with these grades, and in fact may be determined to feeding due to packing the contents of the bin.

Alum dust is present in the ground grade and will cause minor irritation of eyes and respiratory tract. A respirator will protect against alum dust. Alum dust should be thoroughly flushed from the eyes immediately and washed from the skin with water. Gloves should be worn to protect the hands. Because of minor irritation in handling and the possibility of alum dust causing rusting of adjacent machinery, dust removal equipment is desirable.

10.1.1.3 Storage

A typical storage and feeding system for dry alum is shown in Figure 10-1. Bulk alum can be stored in mild steel or concrete bins with dust collector vents located in, above, or adjacent to the equipment room. Recommended storage capacity is about 30 days. Dry alum in bulk can be transferred with screw conveyors, pneumatic conveyors, or bucket elevators made of mild steel. Pneumatic conveyor elbows should have a reinforced backing to withstand abrasion.

Bags and drums of alum should be stored in a dry location. Bag or drum loaded hoppers should have storage capacity for eight hours at the nominal maximum feed rate so personnel are not required to charge the hopper more than once per shift. Converging hopper sections should have a minimum wall slope of 60° to prevent arching.

Bulk Storage hoppers should have a discharge bin gate so feeding equipment may be isolated for servicing. The bin gate should be followed by a flexible connection, and transition hopper chute or hopper which acts as a conditioning chamber over the feeder.

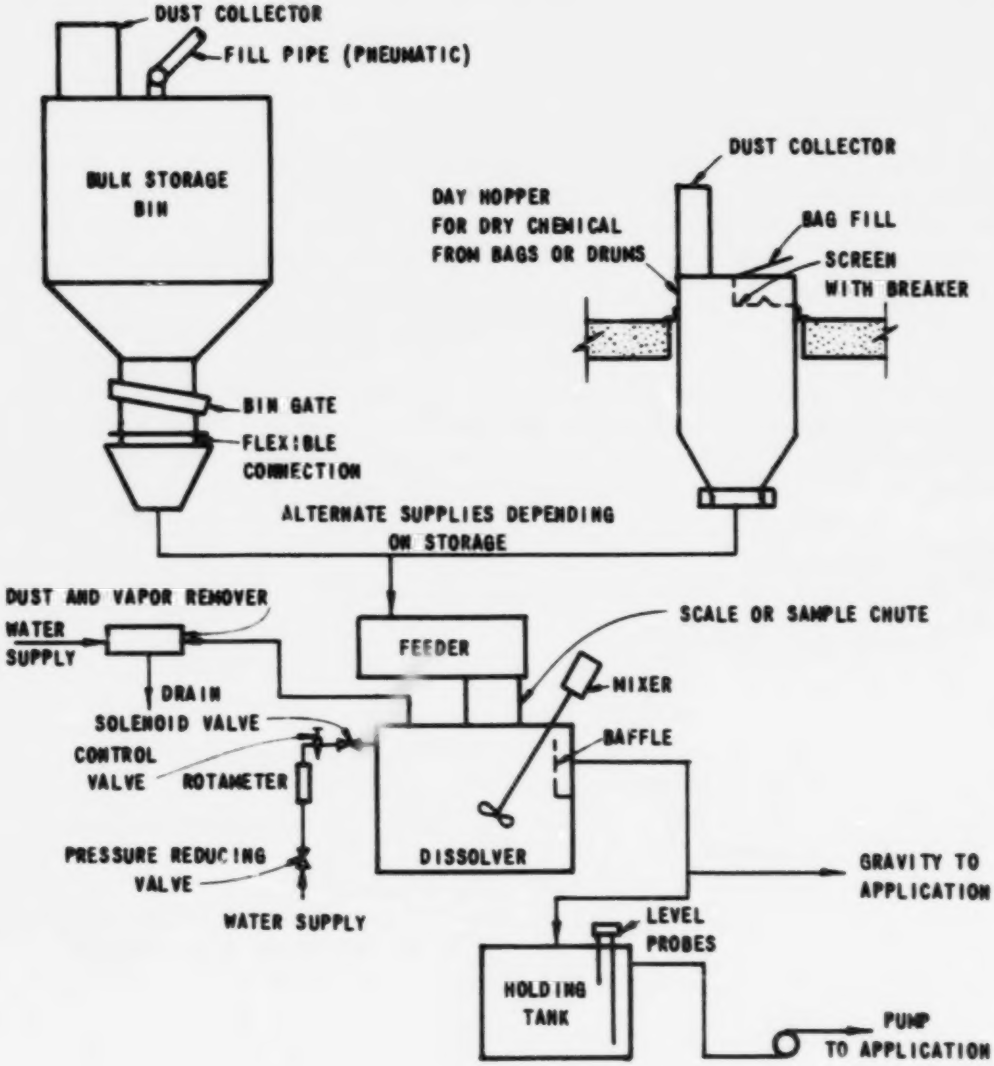
10.1.1.4 Feeding Equipment

The feed system includes all components required for preparation of the chemical solution. Capacities and assemblies should be selected to fulfill individual system requirements. Three basic types of chemical feed equipment are used: volumetric, belt gravimetric, and loss-in-weight gravimetric. Volumetric feeders are usually used where low initial cost and lower delivery capacities are the basis of selection. Volumetric feeder mechanisms are usually exposed to corrosive vapors from the dissolving chamber. Manufacturers usually control this problem by use of an electric heater to keep the feeder housing dry or by using plastic components in the exposed areas.

Volumetric dry feeders are generally of the screw type. Two designs of screw feed mechanism are available. Both allow even withdrawal across the bottom of the feeder hopper to prevent dead zones. One screw design is the variable pitch type with the pitch expanding evenly to the discharge point. The second screw design is the constant pitch-reciprocating type. This type has each half of the screw turned in opposite directions so the turning and reciprocating motion alternately fills one half of the screw while the other half is discharging. The variable pitch screw has one point of discharge, while the constant pitch-reciprocating screw has two points of discharge, one at each end of the

FIGURE 10-1

TYPICAL DRY FEED SYSTEM


DUST COLLECTOR
FILL PIPE (PNEUMATIC)
BULK STORAGE BIN
BIN GATE
FLEXIBLE CONNECTION
DUST COLLECTOR
DAY HOPPER FOR DRY CHEMICAL FROM BAGS OR DRUMS
BAG FILL
SCREEN WITH BREAKER
ALTERNATE SUPPLIES DEPENDING ON STORAGE
DUST AND VAPOR REMOVER
WATER SUPPLY
DRAIN
FEEDER
SCALE OR SAMPLE CHUTE
MIXER
SOLENOID VALVE
CONTROL VALVE
ROTAMETER
BAFFLE
DISSOLVER
PRESSURE REDUCING VALVE
WATER SUPPLY
GRAVITY TO APPLICATION
LEVEL PROBES
HOLDING TANK
PUMP TO APPLICATION

screw. The accuracy of volumetric feeders is influenced by the character of material being fed and ranges between 1% for free-flowing materials to 7% for cohesive materials. This accuracy is volumetric and is not related to accuracy by weight (gravimetric).

Where the greatest accuracy and more economical use of chemical is desired, the loss-in-weight type feeder should be selected. This feeder is suited to low and medium feed rates with a maximum of approximately 4,000 lb/hour. The unit consists of a material hopper and feeding mechanism mounted on enclosed scales. The feed rate controller retracts scale poise weight to deliver the dry chemical at the desired rate. The feeding mechanism must feed at this rate to maintain the balance of the scale. Any unbalance of the scale beam causes a corrective change in output. Continuous comparison of actual hopper weight with set hopper weight prevents cumulative errors. Accuracy of the loss-in-weight feeder is near 1% by weight of the set rate.

Belt-type gravimetric feeders span the capacity ranges of volumetric and loss-in-weight feeders and can be sized for most applications in wastewater treatment. Initial expense falls between that for the volumetric feeder and the loss-in-weight feeder. Belt-type gravimetric feeders consist of a basic belt feeder incorporating a weighing and control system. Feed rates can be varied by changing either the weight per foot of belt, belt speed, or both. Controllers in general use are mechanical, pneumatic, electric and mechanical-vibrating. Accuracy specified for belt-type gravimetric feeders should be within 1% of set rate. This equipment normally includes mild steel hoppers, stainless steel mechanism components, and rubber surfaced feed belts.

Because alum solution is corrosive, dissolving or solution chambers should be constructed of type 316 stainless steel, fiberglass reinforced plastic (FRP), and plastics. Dissolvers should be sized for preparation of the desired solution strength. The most dilute solution strength usually recommended is 0.5 lb of alum to 1 gallon of water, or a 6% solution. The dissolving chamber is designed for a minimum detention time of 5 minutes at the maximum feed rate. Because excessive dilution may be detrimental to coagulation, eductors, or float valves that would ordinarily be used ahead of centrifugal pumps, are not recommended. Dissolvers should be equipped with water meters and mechanical mixers so the water to alum ratio may be properly established and controlled.

10.1.1.5 Piping and Accessories

Pipe made from FRP or plastics (polyvinyl chloride, polyethylene, polypropylene, and similar materials) is recommended for alum solution. Care must be taken to provide adequate support for these piping systems, with close attention given to spans between supports so objectionable deflection will not occur. Lined steel pipe is generally tougher and more rigid, but the cost of providing a near perfect lining may detract from its suitability.

Solution flow by gravity to the point of discharge is desirable. When gravity flow is not possible, transfer components should require little or no dilution. When metering pumps or proportioning weir tanks are used, return of excess flow to a holding tank should be considered. Metering pumps are discussed in the section on liquid alum.

Valves used in solution lines should be plastic, type 316 stainless steel or rubber-lined iron or steel.

10.1.1.6 Pacing and Control

Volumetric and gravimetric feeders are usually adaptable to operation from any standard instrument control and pacing signals.

When solution must be pumped, consideration should be given to use of holding tanks between the dry feed system and feed pumps; solution water supply should be controlled to prevent excessive dilution. The dry feeders may be started and stopped from tank level probes. Variable control metering pumps can transfer alum stock solution to the point of application without further dilution.

Means should be provided for calibration of chemical feeders. Volumetric feeders may be mounted on platform scales. Belt feeders should include a sample chute and box to check actual delivery with set delivery.

Gravimetric feeders are usually furnished with totalizers only. Remote instrumentation is frequently used with gravimetric equipment, but seldom used with volumetric equipment.

10.1.2 Liquid Alum

10.1.2.1 Properties and Availability

Liquid alum is shipped in insulated tank cars or trucks. During the winter it is heated prior to shipment so crystallization will not occur during transit. Liquid alum is shipped at a solution strength of about 4.37% as aluminum or about 8.3% as Al_2O_3 or 49% as $Al_2(SO_4)_3 \quad 14H_2O$. This solution has a density of 11.1 lb/gal at 60°F and contains about 5.4 lb dry alum (17% Al_2O_3) per gallon of liquid. This solution will begin to crystallize at 30°F and crystallizes at 18°F.

Crystallization temperatures of other solution strengths are as follows (1) in Table 10-3:

TABLE 10-3

CRYSTALLIZATION TEMPERATURES OF LIQUID ALUM

Al_2O_3 percent	Temperature of Crystallization °F
5.19	26
6.42	21
6.67	19
6.91	17
7.16	15
7.40	13
7.66	12
7.92	14
8.19	17
8.46	20
8.74	28

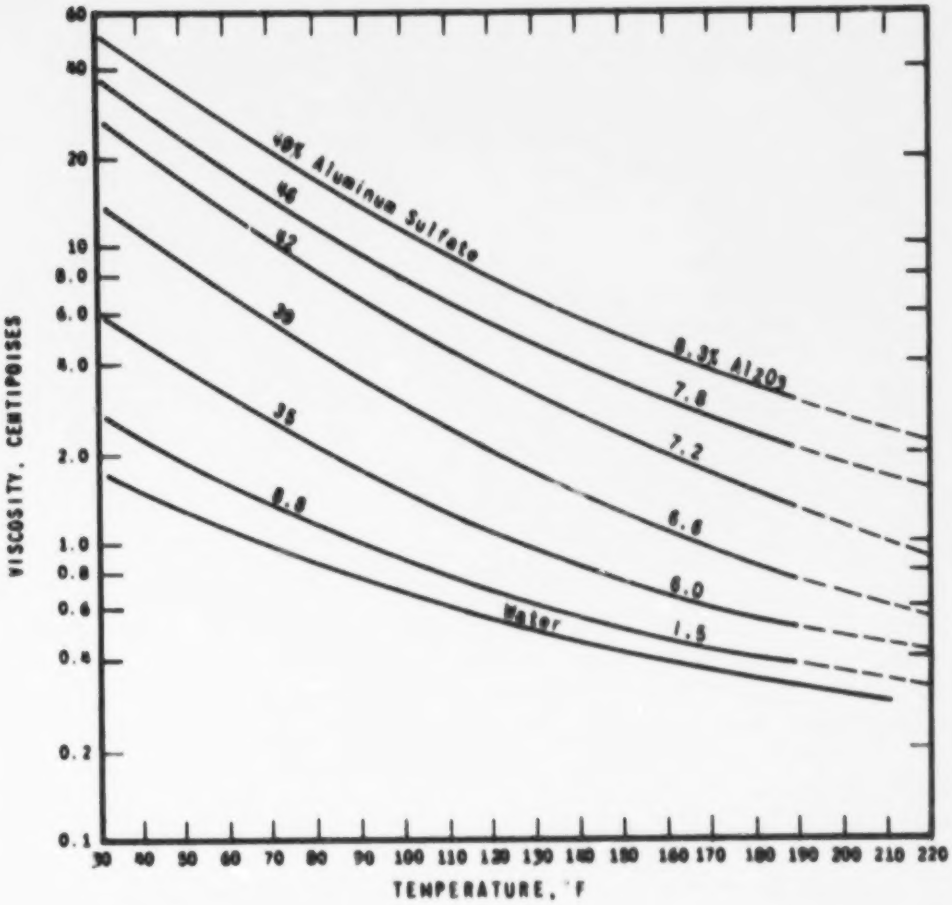
The viscosity of various alum solutions is given in Figure 10-2.

Tank truck lots of 3,000 to 5,000 gallons are available. Tank car lots are available in quantities of 7,000 to 18,000 gallons. The 1974 price of liquid alum on an equivalent dry alum (17% Al_2O_3) basis is about $60 per ton, F.O.B. the point of manufacture. Since liquid alum is an intermediate compound in production of dry alum, the liquid form costs less. However, liquid alum costs more to transport since it is nearly half water, by weight. Therefore, a cost tradeoff point can be established when exact chemical costs and local freight rates are determined. Liquid alum will generally be more economical than dry alum if the point of use is within 100 miles of the manufacturing plant; ease of handling, storing and feeding liquid alum extend its practical transport limit to 200 miles or more.

FIGURE 10-2

VISCOSITY OF ALUM SOLUTIONS

(COURTESY OF ALLIED CHEMICAL CO.)



10.1.2.2 General Design Considerations

Bulk unloading facilities usually must be provided at the treatment plant. Rail cars are constructed for top unloading and therefore require an air supply system and flexible connectors to pneumatically displace the alum from the car. U.S. Department of Transportation regulations concerning chemical tank car unloading should be observed. Tank truck unloading is usually accomplished by gravity or by a truck mounted pump.

Established practice in the treatment field has been to dilute liquid alum prior to application. However, recent studies have shown that feeding undiluted liquid alum results in better coagulation and settling. This is reportedly due to prevention of hydrolysis of the alum.

No particular industrial hazards are encountered in handling liquid alum. However, a face shield and gloves should be worn around leaking equipment. The eyes or skin should be flushed and washed upon contact with liquid alum. Liquid alum becomes very slick upon evaporation and therefore spillage should be avoided.

10.1.2.3 Storage

Liquid alum is stored without dilution at the shipping concentration. Storage tanks may be open if indoors but must be closed and vented if outdoors. Outdoor tanks should also be heated, if necessary, to keep the temperature above 25°F to prevent crystallization. Storage tanks should be constructed of type 316 stainless steel; FRP, steel lined with rubber, polyvinyl chloride, or lead; see subsequent section for further details. Liquid alum can be stored indefinitely without deterioration.

Storage tanks should be sized according to maximum feed rate, shipping time required, and quantity of shipment. Tanks should generally be sized for 1-1/2 times the quantity of shipments. A 10-day to 2-week supply should be provided to allow for unforeseen shipping delays.

10.1.2.4 Feeding Equipment

Various types of gravity or pressure feeding and metering units are available. Figures 10-3 and 10-4 illustrate commonly used feed systems. The rotodip-type feeder or rotameter is often used for gravity feed and the metering pump for pressure feed systems.

The pressure or head available at the point of application frequently determines the feeding system to be used. The rotodip feeder can be supplied from overhead storage by gravity with the use of an internal level control valve, as shown by Figure 10-3. It may also

FIGURE 10-3

ALTERNATIVE LIQUID FEED SYSTEMS

FOR OVERHEAD STORAGE

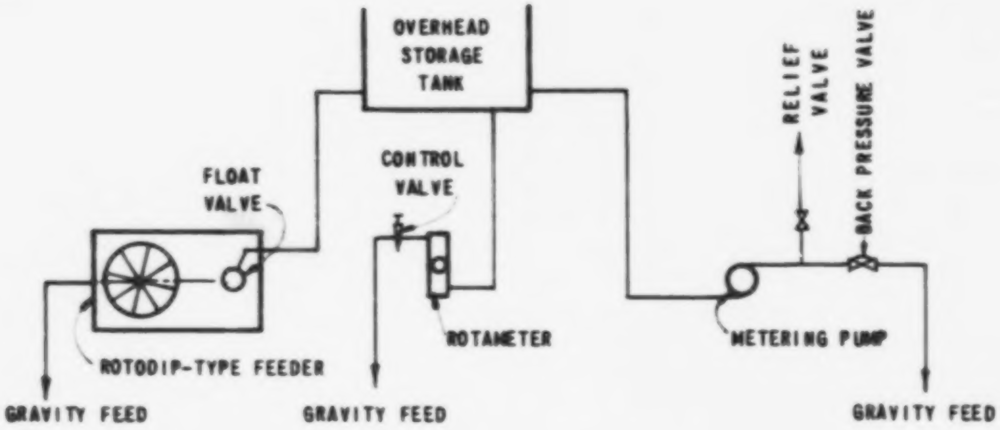
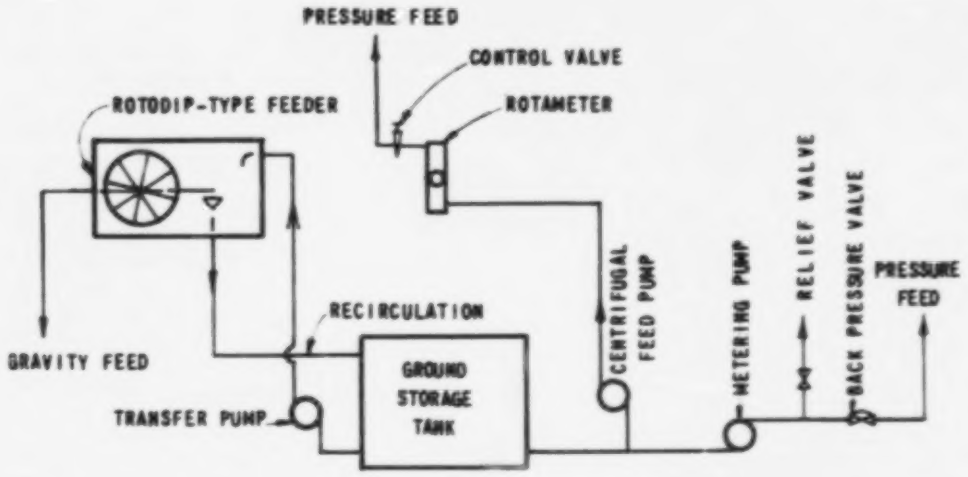

OVERHEAD STORAGE TANK
FLOAT VALVE
CONTROL VALVE
RELIEF VALVE
BACK PRESSURE VALVE
ROTAMETER
METERING PUMP
ROTODIP-TYPE FEEDER
GRAVITY FEED
GRAVITY FEED
GRAVITY FEED

FIGURE 10-4

ALTERNATIVE LIQUID FEED SYSTEMS

FOR GROUND STORAGE


PRESSURE FEED
CONTROL VALVE
ROTODIP-TYPE FEEDER
ROTAMETER
RELIEF VALVE
BACK PRESSURE VALVE
CENTRIFUGAL FEED PUMP
METERING PUMP
PRESSURE FEED
RECIRCULATION
GRAVITY FEED
GROUND STORAGE TANK
TRANSFER PUMP

be supplied by a centrifugal pump. The latter arrangement requires an excess flow return line to the storage tank, as shown by Figure 10-4. Centrifugal pumps should be direct-connected but not close-coupled because of possible leakage into the motor, and should be constructed of type 316 stainless steel, FRP, and plastics.

Metering pumps, currently available, allow a wide range of capacity to compare with the rotodip and rotameter systems. Hydraulic diaphragm type pumps are preferable to other type pumps and should be protected with an internal or external relief valve. A back pressure valve is usually required in the pump discharge to provide efficient check valve action. Materials of construction for feeding equipment should be as recommended by the manufacturer for the service, but depending on the type of system, will generally include type 316 stainless steel, FRP, plastics, and rubber.

10.2.1.5 Piping and Accessories

Piping systems for alum should be FRP, plastics (subject to temperature limits), type 316 stainless steel, or lead. Piping and valves used for alum solutions are also discussed in the preceding section on dry alum.

10.1.2.6 Pacing and Control

The feeding systems described above are volumetric, and the feeders generally available can be adapted to receive standard instrument pacing signals. The signals can be used to vary motor speed, variable speed transmission setting, stroke speed and stroke length where applicable. A totalizer is usually furnished with a rotodip-type feeder, and remote instruments are available. Instrumentation is rarely used with rotameters and metering pumps.

10.1.3 Dry Sodium Aluminate

10.1.3.1 Properties and Availability

Dry sodium aluminate, $Na_2Al_2O_4$, is available from manufacturers listed in a subsequent section. Dry sodium aluminate is shipped in 50 lb bags and has a bulk density ranging from 40 to 50 lb/ft^3. The Al_2O_3 content ranges from 41 to 46%. Dry sodium aluminate is noncorrosive and the pH of a 1% solution is about 11.9. Manufacturers should be consulted for more precise specifications of their product.

The 1974 price of dry sodium aluminate in 40,000 lb shipments ranges from $0.13 to $0.15 per lb, F.O.B. the point of manufacture. Prices increase substantially for smaller shipments.

10.1.3.2 General Design Considerations

Requirements for dry sodium aluminate feed systems are generally similar to those for dry aluminum sulfate. Dry sodium aluminate is not available in bulk quantities. Therefore, the small, day-type hoppers with manual filling arrangements as shown by Figure 10-1 are used. Precautionary measures for handling sodium aluminate are similar to those for strong alkalies, such as caustic soda. Contact with skin, eyes, and clothing should be avoided. Aluminate dust or solution spray should not be breathed.

10.1.3.3 Storage

Dry sodium aluminate is stored as received, in bags, and at optimum conditions of 60 to 90°F; the recommended storage limit is 6 months. Hopper material of mild steel is completely adequate. This chemical may or may not be free flowing, depending on the manufacturer and grade used. Therefore, hopper agitation may be required. Sodium aluminate deteriorates on exposure to the atmosphere and care should be taken to avoid tearing of bags.

10.1.3.4 Feeding Equipment

Materials of construction for dissolving chambers may be mild steel or stainless steel and selection may be influenced by conformity with adjacent equipment. Equipment similar to that shown by Figure 10-1 is applicable. Standard practice for the free-flowing grade of sodium aluminate calls for dissolvers sized for 0.5 lb per gallon or 6% solution strength with a dissolver detention time of 5 minutes at the maximum feed rate.

After dissolving dry sodium aluminate in the preparation of batch solutions, agitation should be minimized or eliminated to prevent deterioration of the solution. Air agitation is not recommended, and solution tanks should be covered to prevent carbonation of the solution.

10.1.3.5 Piping and Accessories

Materials for piping and transporting dry sodium aluminate solution may be mild steel, iron, type 304 stainless steel, concrete, or plastics. The use of copper, copper alloys, and rubber should be avoided.

10.1.3.6 Pacing and Control

Pacing and control fundamentals are similar to those described for dry alum. The amount of dilution is not a consideration in the use of sodium aluminate. Therefore, the use of float valves to satisfy centrifugal pump suction, and the use of eductors, is permissible.

10.1.4 Liquid Sodium Aluminate

10.1.4.1 Properties and Availability

Liquid sodium aluminate is available from the manufacturers listed in a subsequent section. There is considerable variety in the composition of liquid sodium aluminate from the manufacturers listed. The Al_2O_3 content varies from 4.9 to 26.7%. The lower solution strengths are usually more expensive because of the cost of freighting the solution water. Because of the variety of solution strengths available, the manufacturers should be contacted for more specific information on density, viscosity, and cost.

Liquid sodium aluminate is available in 30-gallon drums (380 lb), tank truck, and tank car quantities. The approximate 1974 price of liquid sodium aluminate in 40,000 lb quantities is about $0.05 per lb ($Al_2O_3$ content, 26.4% or 19.9%) F.O. B. point of manufacture.

10.1.4.2 General Design Considerations

Because of the alkaline nature of sodium aluminate, it should not be used in contact with brass, copper, aluminum or rubber. Liquid sodium aluminate is a strong alkali and the same precautions should be exercised in handling it as in handling caustic soda.

10.1.4.3 Storage

Liquid sodium aluminate is usually stored at the shipping concentration, either in the shipping drums or in mild steel tanks. Storage tanks may be located indoors or outdoors; however, outdoor tanks should be provided with facilities for indirect heating. The maximum recommended length of storage is two to three months. Bulk shipments can be unloaded by gravity, pumping, or air pressure. However, if air is used, it should first be passed through lime-caustic soda breathers to remove carbon dioxide. Steam injection facilities are required at the unloading site.

10.1.4.4 Feeding Equipment

Feeding equipment and systems as described for liquid alum generally apply to liquid sodium aluminate except with changes of requirements regarding dilution and materials of construction as described above.

Liquid sodium aluminate may be fed at shipping strength or diluted to a stable 5 to 10% solution. Stable solutions are prepared by direct addition of low hardness water and mild agitation. Air agitation is not recommended.

10.1.4.5 Piping and Accessories

Piping requirements are the same as previously indicated for solutions of dry aluminate.

10.1.4.6 Pacing and Control

System pacing and control requirements are the same as described for liquid alum.

10.1.5 Aluminum Chloride

Aluminum chloride is not as readily available as other compounds previously discussed, and in some cases it does not enjoy a competitive price structure; nevertheless, it should be considered as a source of aluminum by designers undertaking phosphorus precipitation by metal addition.

The majority of the aluminum chloride production facilities in this country are located near petroleum refining centers and petrochemical complexes. This is because those industries consume three-fourths of the total production. Miscellaneous users of the balance of the overall production include makers of antiperspirants and a variety of other unrelated manufacturing operations. Therefore, a reliable and economical source of aluminum chloride may, in fact, be near the proposed wastewater treatment project.

Much of the aluminum chloride produced is the form of semi-pure anhydrous crystals with an off-white color. This product is derived from direct chlorination of scrap aluminum, and has a molecular weight of 133.3 and purities near 99 percent. In some instances purities may drop to 96 percent which is satisfactory for wastewater treatment if analysis shows other compounds present are acceptable.

Another solid form is produced by crystallization from hydrochloric acid in the form of a hexahydrate, in which six molecules of water attach to each molecule of aluminum chloride.

The most usual commercial form of liquid aluminum chloride is in a 32-degree Baume solution; it contains 28 percent aluminum chloride by weight. This particular liquor is not universally produced, however, so designers should be prepared to consider other concentrations with a fairly wide variety of other chemical compounds present.

Containers for shipping and storing should be similar to those used for dry and liquid alum. Handling and feeding requirements are also similar to those used with alum. Designers should check with local producers for further details.

Designers may contact manufacturers operating in at least seventeen plants in the U.S. shown in Table 10-4:

TABLE 10-4

SUPPLIERS OF ALUMINUM CHLORIDE

Allied Chemical Corp.:	El Segundo, Calif.; Hegewisch, Ill.; Elberta, N.Y.
Clinton Chemical Co.:	Houston, Texas
Dow Chemical Co.:	Freeport, Texas
E. I. du Pont de Nemours & Co., Inc.:	Grasselli, N.J.; East Chicago, Ind.
FMC Corporation:	Nitro, W. Va.
Hercules Alchlor Chemical Co., Inc.:	Ravenna, Ohio
Monsanto Co.:	Everett, Mass.; Texas City, Texas
Pearsall Chemical Corp.:	Phillipsburg, N.J.; La Porte, Texas
Reheis Co., Inc.:	Berkeley Heights, N.J.
Stauffer Chemical Co.:	Dominguez, Calif.; Baton Rouge, La.; Elkton, Md.

10.1.6 Estimated Initial Costs of Adding Liquid Alum Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of chemical storage and feeding equipment for 1, 10, and 100 mgd plants have been prepared based on the use of liquid alum (8.3% Al_2O_3) for phosphorus precipitation. A mole ratio of Al:P of 2:1 was assumed to treat an influent phosphorus concentration of 10 mg/l as P. This corresponds to an alum dose of 191 mg/l (17.4 mg/l as aluminum).

Chemical feed equipment was sized for a peak feed rate of twice that calculated from the mole ratio. Storage was provided for at least 15 days at the average feed rate. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the source of chemical supply. Piping and buildings to house the feeding equipment are not included in the estimates. At many locations the chemical feeding equipment can be installed in existing buildings. In all estimates, the costs include an allowance for the contractor's installation, overhead, and profit, and an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are shown in Table 10-5. These figures correspond to an EPA Construction Cost Index of 185:

TABLE 10-5

COST OF LIQUID ALUM FEED FACILITIES

Plant Size (mgd)	Estimated 1973 Cost ($)
1	15,000
10	50,000
100	600,000

For the 1.0 mgd plant, feeding equipment for liquid alum includes two 25 gph hydraulic diaphragm pumps (one operating and one standby) with the necessary accessories and equipment to pace the feed rate with plant flow. Two 3,000-gallon FRP tanks with accessories are provided for storage of liquid alum. The total capacity of 6,000 gallons allows some flexibility in ordering and receiving 4,000-gallon tank truck shipments.

For the 10 mgd plant, liquid alum is fed by three 125 gph rotodip-type feeders (two operating and one standby) with the necessary accessories and control panel for proportioning chemical feed to flow. Alum is stored in four 11,500-gallon FRP tanks.

For the 100 mgd plant, liquid alum is fed with seven 415 gph rotodip-type feeders (six operating and one standby) with the necessary control equipment. Storage costs are based on ten 50,000-gallon underground concrete tanks for the rotodip feeders but eliminates the need for heating the tanks. Three 500-gallon FRP day tanks are included in the estimate; one for each pair of feeders. Four alum transfer pumps (three operating and one standby), with a capacity of 50 gpm at 50 ft of head are provided.

The cost of feeding and storage facilities for the 100 mgd plant is greater than 10 times the cost of facilities for the 10 mgd plant. This is because underground storage facilities were the basis of design for the 100 mgd plant in contrast to FRP tanks located in an existing building for the 10 mgd plant. The cost of a building for the 1.0 and 10.0 mgd plants would raise the cost substantially. FRP tanks could be used for the larger plant; however, structural design may necessitate special construction requirements for underground FRP tanks. FRP tanks located above ground and on concrete foundations would require special insulation and heating.

10.2 Iron Compounds

10.2.1 Liquid Ferric Chloride

10.2.1.1 Properties and Availability

Liquid ferric chloride (2) is a corrosive, dark brown oily-appearing solution having a typical unit weight between 11.2 and 12.4 lb/gal (35 to 45% $FeCl_3$). The ferric chloride content of these solutions is 3.95 to 5.58 lb/gal. Shipping concentrations vary from summer to winter due to the relatively high crystallization temperature of the more concentrated solutions as shown by Figure 10-5. The pH of a 1% solution is 2.0.

The molecular weight of ferric chloride is 162.22. Viscosities of ferric chloride solutions at various temperatures are presented in Figure 10-6.

Liquid ferric chloride is shipped in 3,000 to 4,000 gallon bulk truckload lots, in 4,000 to 10,000 gallon bulk carload lots, and in 5 and 13 gallon carboys. Liquid ferric chloride is produced at the locations listed in a subsequent section. The 1974 price of liquid ferric chloride in bulk quantities is $0.045 per lb (as ferric chloride), F.O.B. point of manufacture.

FIGURE 10-5

FREEZING POINT CURVE FOR COMMERCIAL

FERRIC CHLORIDE SOLUTIONS

(ADAPTED FROM DOW CHEMICAL CO.)

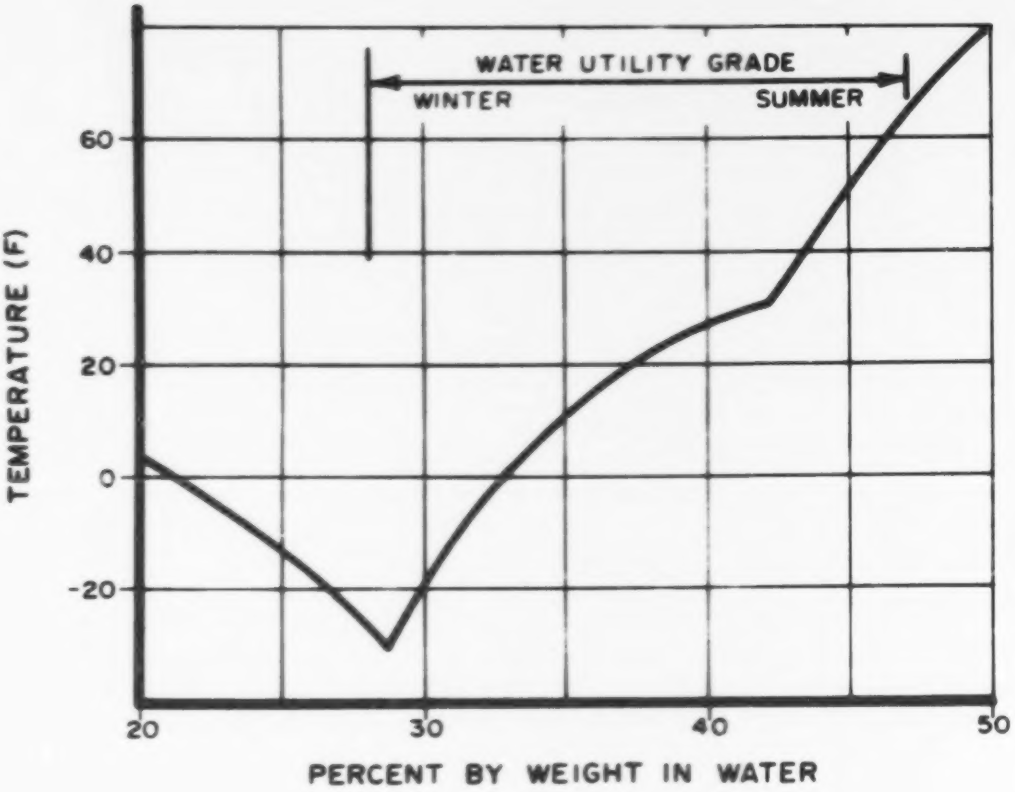
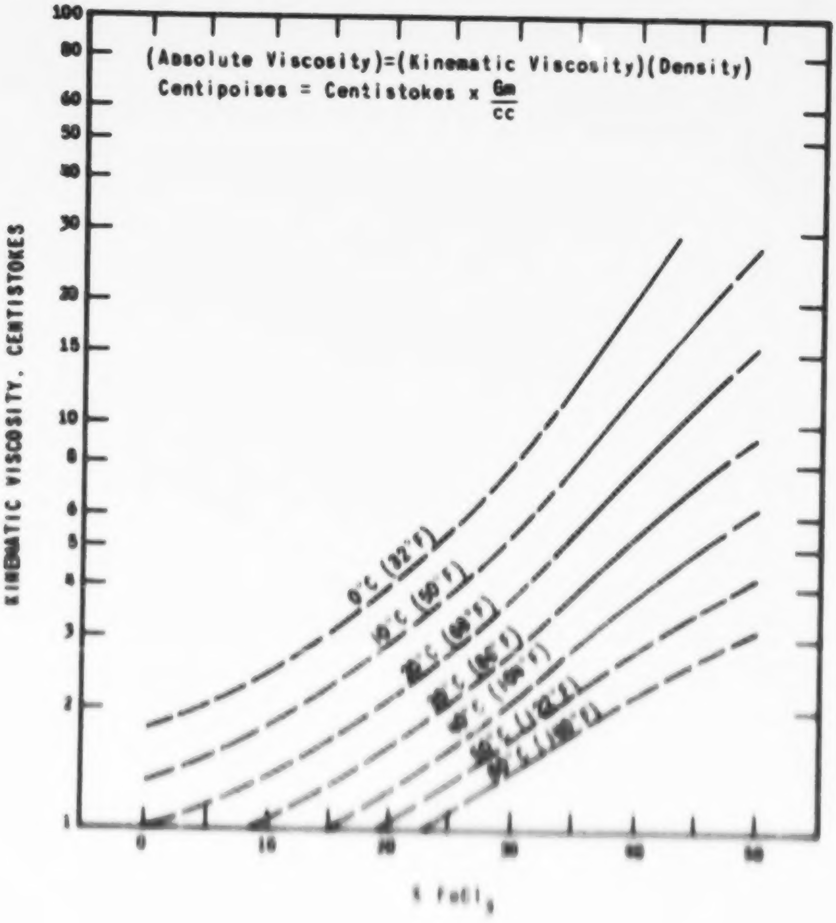


FIGURE 10-6

VISCOSITY VS COMPOSITION OF FERRIC CHLORIDE SOLUTIONS AT VARIOUS TEMPERATURES

(COURTESY OF DOW CHEMICAL CO.)



Tank trucks and cars are normally unloaded pneumatically, and operating procedures must be closely followed to avoid spills and accidents. The safety vent cap and assembly (painted red) should be removed prior to opening the unloading connection to depressurize the tank car or truck, prior to unloading.

10.2.1.2 General Design Considerations

Ferric chloride solutions are corrosive to many common materials and cause stains which are difficult to remove. Areas which are subject to staining should be protected with resistant paint or rubber mats.

Normal precautions should be employed when cleaning ferric chloride handling equipment. Workmen should wear rubber gloves, rubber apron, and goggles or a face shield. If ferric chloride comes in contact with the eyes or skin, flush with copious quantities of running water and call a physician. If ferric chloride is ingested, induce vomiting and call a physician.

10.2.1.3 Storage

Ferric chloride solution can be stored as shipped. Storage tanks should have a free vent or vacuum relief valve. Tanks may be constructed of FRP, rubber lined steel, or plastic lined steel. Resin impregnated carbon or graphite are also suitable materials for storage containers.

It may be necessary in most instances to house liquid ferric chloride tanks in heated areas or provide tank heaters or insulation to prevent crystallization. Ferric chloride can be stored for long periods of time without deterioration. The total storage capacity should be 1-1/2 times the largest anticipated shipment, and should provide at least a 10-day to 2-week supply of the chemical at the design average dose.

10.2.1.4 Feeding Equipment

Feeding equipment and systems described for liquid alum generally apply to ferric chloride except for materials of construction, and the use of glass tube rotameters.

It may not be desirable to dilute the ferric chloride solution from its shipping concentration to a weaker feed solution because of possible hydrolysis. Ferric chloride solutions may be transferred from underground storage to day tanks with impervious graphite or rubber lined self-priming centrifugal pumps having teflon rotary and stationary seals. Because of the tendency for liquid ferric chloride to stain or deposit, glass-tube rotameters should not be used for metering this solution. Rotodip feeders and diaphragm metering pumps are often used for ferric chloride, and should be constructed of materials such as rubber lined steel and plastics.

10.2.1.5 Piping and Accessories

Materials for piping and transporting ferric chloride should be rubber or Saran lined steel, hard rubber, FRP, or plastics. Valving should consist of rubber or resin lined diaphragm valves, Saran lined valves with teflon diaphragms, rubber sleeved pinch type valves, or plastic ball valves. Gasket material for large openings such as manholes in storage tanks should be soft rubber; all other gaskets should be graphite-impregnated blue asbestos, teflon, or vinyl.

10.2.1.6 Pacing and Control

System pacing and control requirements are similar to those discussed previously for liquid alum.

10.2.2 Ferrous Chloride (Waste Pickle Liquor)

10.2.2.1 Properties and Availability

Ferrous chloride, $FeCl_2$, as a liquid is available in the form of waste pickle liquor from steel processing. The liquor weighs between 9.9 and 10.4 lb/gal and contains 20 to 25% ferrous chloride or about 10% available iron. A 22% solution of ferrous chloride will crystallize at a temperature of -4°F. The molecular weight of ferrous chloride is 126.76. Free acid in waste pickle liquor can vary from 1 to 10% and usually averages about 1.5 to 1.0%. Ferrous chloride is slightly less corrosive than ferric chloride.

Waste pickle liquor is available in 4,000-gallon truckload lots and a variety of carload lots. In most instances the availability of waste pickle liquor will depend on the proximity to steel processing plants. Dow Chemical Company produces a waste pickle liquor, having a ferrous chloride content of about 22% at a 1974 price of $0.045 per lb of ferrous chloride in bulk car or truckload quantities, F.O.B. Midland, Michigan.

10.2.2.2 General Design Considerations

Since ferrous chloride or waste pickle liquor may not be available on a continuous basis, storage and feeding equipment should be suitable for handling ferric chloride. Therefore, the ferric chloride section should be referred to for storage and handling details.

10.2.3 Ferric Sulfate

10.2.3.1 Properties and Availability

Ferric sulfate, $Fe_2(SO_4)_3 \cdot X\ H_2O$, is marketed as a dry, partially hydrated product with seven water molecules. Typical properties are presented in Table 10-6 below (3):

TABLE 10-6

PROPERTIES OF FERRIC SULFATE

Molecular Weight	526
Bulk Density, lb/cu ft	56-60
Water Soluble Iron Expressed as Fe	21.5%
Water Soluble Fe^{+3}	19.5%
Insolubles Total	2.0%
Free Acid	2.5%
Moisture @ 105°C	2.0%

Ferric sulfate is shipped in car and truckload lots of 50 lb and 100 lb moistureproof paper bags and 200 lb and 400 lb fiber drums. Bulk carload shipments in box and closed hopper cars are available. The major producer is Cities Service Company, with a plant located at Copper Hill, Tennessee.

The 1973 price of bulk ferric sulfate (21.8% Fe) is about $39 per ton, F.O.B. Copper Hill, Tennessee. Bagging adds some $6 to $11 per ton over the bulk rate.

General precautions should be observed when handling ferric sulfate, such as wearing goggles and dust masks, and areas of the body that come in contact with the dust or vapor should be washed promptly.

10.2.3.2 General Design Considerations

Aeration of ferric sulfate should be held to a minimum because of the hygroscopic nature of the material, particularly in damp atmospheres. Mixing of ferric sulfate and quicklime in conveying and dust vent systems should be avoided as caking and excessive heating can result. The presence of ferric sulfate and lime in combination has been known to destroy cloth bags in pneumatic unloading devices (4). Because ferric sulfate in the presence of moisture will stain, precautions similar to those discussed for ferric chloride should be observed.

10.2.3.3 Storage

Ferric sulfate is usually stored in the dry state either in the shipping bags or in bulk in concrete or steel bins. Bulk storage bins should be as tight as possible to avoid moisture absorption, but dust collector vents are permissible and desirable. Hopper walls should slope 36% or more.

Bins may be located inside or outside and the material transferred by bucket elevator, screw or air conveyors. Ferric sulfate stored in bins usually absorbs some moisture and forms a thin protective crust which retards further absorption until the crust is broken.

10.2.3.4 Feeding Equipment

Feed solutions are usually made up at a water to chemical ratio of 2:1 to 8:1 (on a weight basis) with the usual ratio being 4:1 with a 20-minute detention time. Care must be taken not to dilute ferric sulfate solutions to less than 1% to prevent hydrolysis and deposition of ferric hydroxide. Ferric sulfate is actively corrosive in solution, and dissolving and transporting equipment should be fabricated of type 316 stainless steel, rubber, plastics, ceramics or lead.

Dry feeding requirements are similar to those for dry alum except that belt type feeders are rarely used because of their open type of construction. Closed construction, as found in the volumetric and loss-in-weight type feeders, generally exposes a minimum of operating components to the vapor, and thereby minimizes maintenance. A water jet vapor remover should be provided at the dissolver to protect both the machinery and operator.

10.2.3.5 Piping and Accessories

Piping systems for ferric sulfate should be FRP, plastics, type 316 stainless steel, rubber or glass.

10.2.3.6 Pacing and Control

System pacing and control are the same as discussed for dry alum.

10.2.4 Ferrous Sulfate

10.2.4.1 Properties and Availability

Ferrous sulfate or copperas is a byproduct of pickling steel and is produced as granules, crystals, powder, and lumps. The most common commercial form of ferrous sulfate is $FeSO_4 \cdot 7H_2O$, with a molecular weight of 278, and containing 55 to 58% $FeSO_4$ and 20 to 21% Fe. The product has a bulk density of 62 to 66 lb/ft^3. When dissolved, ferrous sulfate is acidic. The composition of ferrous sulfate may be quite variable and should be established by consulting the nearest manufacturers.

Bulk, drum (400-lb) and bag (50 and 100 lb) shipments are available from producers at locations listed in a subsequent section. The 1974 bulk price of ferrous sulfate in bulk carload and truckload quantities is about $28 per ton (21% Fe), with an additional $8 per ton for bagging.

Ferrous sulfate is also available in a wet state in bulk form from some plants. This form is likely to be difficult to handle and the manufacturer should be consulted for specific information and instructions.

Dry ferrous sulfate cakes at storage temperatures above 68°F, is efflorescent in dry air, and oxidizes and hydrates further in moist air.

General precautions similar to those for ferric sulfate with respect to dust and handling acidic solutions, should be observed when working with ferrous sulfate. Mixing quicklime and ferrous sulfate produces high temperatures and the possibility of fire.

10.2.4.2 General Design Considerations

The granular form of ferrous sulfate has the best feeding characteristics and gravimetric or volumetric feeding equipment may be used.

The optimum chemical to water ratio for continuous dissolving is 0.5 lb/gal or 6% with a detention time of 5 minutes in the dissolvers. Mechanical agitation should be provided in the dissolver to assure complete solution. Lead, rubber, iron, plastics, and type 304 stainless steel can be used as construction materials for handling solutions of ferrous sulfate.

Storage, feeding and transporting systems probably should be suitable for handling ferric sulfate as an alternative to ferrous sulfate.

10.2.5 Estimated Initial Costs of Adding Liquid Ferric Chloride Feed Systems to Existing Wastewater Treatment Plants

Costs of chemical storage and feeding equipment for 1, 10, and 100 mgd plants have been estimated based on the use of liquid ferric chloride (basis: 35% ferric chloride) for phosphorus precipitation. A mole ratio of Fe:P of 2.0:1 was assumed to treat an influent phosphorus concentration of 10 mg/l as P. This corresponds to a ferric chloride dose of 104 mg/l (36 mg/l as Fe^{3+}).

Chemical feed equipment was sized for a peak feed rate of twice that calculated from the mole ratio. Storage was provided for at least 15 days at the average feed rate. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the souce of chemical supply. Piping and buildings to house the feeding equipment are not included. At many locations the chemical feeding equipment can be installed in existing buildings. All estimated costs include an allowance for the contractor's installation, overhead, and profit plus an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are shown in Table 10-7. These figures correspond to an EPA Construction Cost Index of 185.

TABLE 10-7

COSTS OF LIQUID FERRIC CHLORIDE FEED FACILITIES

Plant Size (mgd)	Estimated 1973 Cost ($)
1	15,000
10	50,000
100	500,000

For the 1 mgd plant, the feeding equipment for liquid ferric chloride includes two 18 gph, hydraulic diaphragm pumps (one operating and one standby) with the necessary accessories and equipment to pace the feed rate to the plant flow. Liquid ferric chloride is stored in two 3,000-gallon FRP tanks with accessories. The 6,000-gallon total capacity allows use of liquid ferric chloride in 4,000-gallon tank truck quantities.

For the 10 mgd plant, liquid ferric chloride is fed by three 90 gph rotodip-type feeders (two operating and one standby) with the necessary accessories and control panel for proportioning chemical feed to flow.

For the 100 mgd plant, seven 305 gph rotodip-type feeders (six operating and one standby) are provided along with the necessary control equipment. Storage is provided in eight 50,000-gallon rubber-lined concrete tanks located underground. The underground storage necessitates transfer pumps and day tanks for the rotodip feeders, but eliminates the need for heating the tanks. Three 500-gallon FRP day tanks are included along with four ferric chloride transfer pumps sized for 35 gpm each at 50 ft of head.

With regard to the cost of the feed system for the 100 mgd plant being out of proportion to that for a 10 mgd plant, the reader is referred to Section 10.1.5 of this chapter for the discussion of underground concrete storage vessels as opposed to FRP tanks.

10.3 Lime

10.3.1 Quicklime

10.3.1.1 Properties and Availability

Quicklime, CaO, has a density range of approximately 55 to 75 lb/ft^3, and a molecular weight of 56.08. A slurry for feeding, called milk of lime, can be prepared with up to 45% solids. Lime is only slightly soluble, and both lime dust and slurries are caustic in nature. A saturated solution of lime has a pH of about 12.4.

Lime can be purchased in bulk in both car and truckload lots. It is also shipped in 80 and 100 lb multiwall "moistureproof" paper bags. Lime is produced at the locations listed in a subsequent section (5).

1974 prices for bulk pebble quicklime range from $18 per ton to $22 per ton with the higher prices generally in the far west, and higher than average in the north. Bagging adds approximately $5 per ton to the cost.

The CaO content of commercially available quicklime can vary quite widely over an approximately range of 70 to 96%. Content below 88% is generally considered below standard in the municipal use field (6). Purchase contracts are often based on 90% CaO content with provisions for payment of a bonus for each 1% over and a penalty for each 1% under the standard. A CaO content less than 75% probably should be rejected because of excessive grit and difficulties in slaking.

Workmen should wear protective clothing and goggles to protect the skin and eyes, as lime dust and hot slurry can cause severe burns. Areas contacted by lime should be washed immediately. Lime should not be mixed with chemicals which have water of hydration. The lime will be slaked by the water of hydration causing excessive temperature rise and possibly explosive conditions. Conveyors and bins used for more than one chemical should be thoroughly cleaned before switching chemicals.

10.3.1.2 General Design Considerations

Pebble quicklime, all passing a 3/4-in screen and not more than 5% passing a No. 100 screen, is normally specified because of easier handling and less dust. Hopper agitation is generally not required with the pebble form. Published slaker capacity ratings require "soft or normally burned" limes which provide fast slaking and temperature rise, but poorer grades of limes may also be satisfactorily slaked by selection of the appropriate slaker retention time and capacity.

10.3.1.3 Storage

Storage of bagged lime should be in a dry place, and preferably elevated on pallets to avoid absorption of moisture. System capacities often make the use of bagged quicklime impractical. Maximum storage period is about 60 days.

Bulk lime is stored in air-tight concrete or steel bins having a 60^{0} slope on the bin outlet. Bulk lime can be conveyed by conventional bucket elevators and screw, belt, apron, drag-chain, and bulk conveyors of mild steel construction. Pneumatic conveyors subject the lime to air-slaking and particle sizes may be reduced by attrition. Dust collectors should be provided on manually and pneumatically filled bins.

10.3.1.4 Feeding Equipment

A typical lime storage and feed system is illustrated in Figure 10-7. Quicklime feeders are usually limited to the belt or loss-in-weight gravimetric types because of the wide variation of the bulk density. Feed equipment should have an adjustable feed range of at least 20:1 to match the operating range of the associated slaker. The feeders should have an over-under feed rate alarm to immediately warn of operation beyond set limits of control. The feeder drive should be instrumented to be interrupted in the event of excessive temperature in the slaker compartment.

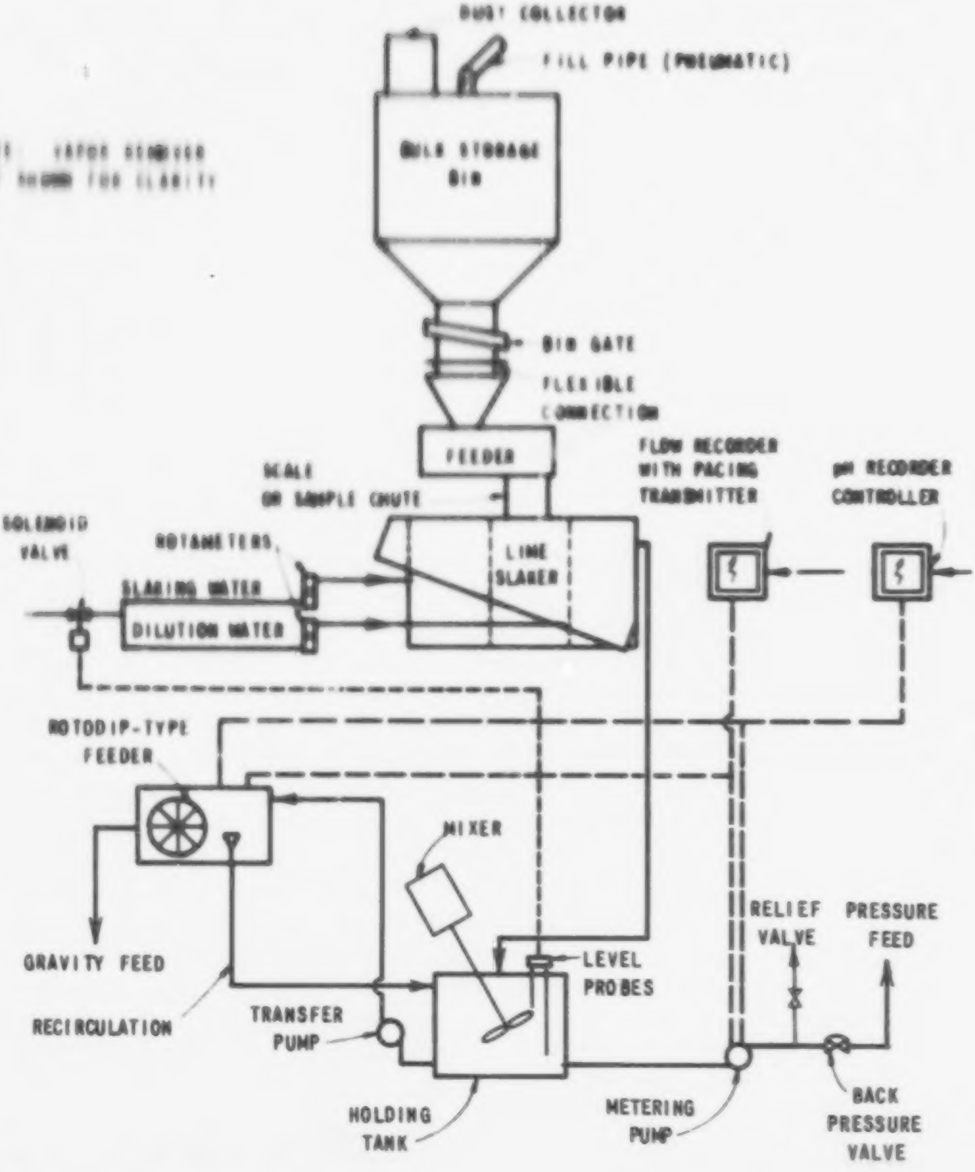
Lime slakers for wastewater treatment should be of the continuous type, and the major components should include one or more slaking compartments, a dilution compartment, a grit separation compartment and a continuous grit remover. Commercial designs vary in regard to the combination of water to lime, slaking temperature, and slaking time, in obtaining the "milk of lime" suspensions.

The "paste" type slaker admits water as required to maintain a desired mixing viscosity. This viscosity therefore sets the operating retention time of the slaker. The paste slaker usually operates with a low water to lime ratio (approximately 2:1 by weight) elevated temperature, and five minute slaking time at maximum capacity.

The "detention" type slaker admits water to maintain a desired ratio with the lime, and therefore the lime feed rate sets the retention time of the slaker. The detention slaker operates with a wide range of water to lime ratios (2.5:1 and 6:1), moderate temperature, and a 10-minute slaking time at maximum capacity. A water to lime ratio of from 3.5:1 to 4:1 is most often used. The operating temperature in lime slakers is a function of the water to lime ratio, lime quality, heat transfer, and water temperature. Lime slaking

FIGURE 10-7

TYPICAL LIME FEED SYSTEM


DUST COLLECTOR
FILL PIPE (PNEUMATIC)
BULK STORAGE BIN
BIN GATE
FLEXIBLE CONNECTION
FEEDER
SCALE OR SAMPLE CHUTE
FLOW RECORDER WITH PACING TRANSMITTER
pH RECORDER CONTROLLER
SOLENOID VALVE
ROTAMETERS
SLAKING WATER
DILUTION WATER
LIME SLAKER
ROTODIP-TYPE FEEDER
MIXER
GRAVITY FEED
RECIRCULATION
TRANSFER PUMP
LEVEL PROBES
HOLDING TANK
RELIEF VALVE
PRESSURE FEED
METERING PUMP
BACK PRESSURE VALVE

NOTE: VAPOR REMOVER NOT SHOWN FOR CLARITY evolves heat in hydrating the CaO to $Ca(OH)_2$ and, therefore, vapor removers are required for feeder protection.

10.3.1.5 Piping and Accessories

Lime slurry should be transported by gravity in open channels wherever possible. Piping channels and accessories may be rubber, iron, steel, concrete, and plastics. Glass tubing, such as that in rotameters, will cloud rapidly and therefore should not be used. Any abrupt directional changes in piping should include plugged tees or crosses to allow rodding-out of deposits. Long sweep elbows should be provided to allow the piping to be cleaned by the use of a cleaning "pig". Daily cleaning is desirable.

Milk of lime transfer pumps should be of the open impeller centrifugal type. Pumps having an iron body and impeller with bronze trim are suitable for this purpose. Rubber lined pumps with rubber covered impellers are also frequently used. Make-up tanks are usually provided ahead of centrifugal pumps to ensure a flooded suction at all times. "Plating-out" of lime is minimized by the use of soft water in the make-up tank. Turbine pumps and eductors should be avoided in transferring milk of lime because of scaling problems.

10.3.1.6 Pacing and Control

Lime slaker water proportioning is integrally controlled or paced from the feeder. Therefore, the feeder-slaker system will follow pacing controls applied to the feeder only. As discussed previously, gravimetric feeders are adaptable to receive most standard instrumentation pacing signals. Systems can be instrumented to allow remote pacing with telemetering of temperature and feed rate to a central panel for control purposes.

The lime feeding system may be controlled by an instrumentation system integrating both plant flow and pH of the wastewater after lime addition. However, it should be recognized that pH probes require daily maintenance in this application to monitor the pH accurately. Deposits tend to build up on the probe and necessitate frequent maintenance. The low pH lime treatment systems (pH 9.5 to 10.0) can be more readily adapted to this method of control than high lime treatment systems (pH 11.0 or greater) because less maintenance of the pH equipment is required. In a closed loop pH-flow control system, milk of lime is prepared on a batch basis and transferred to a holding tank with variable output feeders set by the flow and pH meters to proportion the feed rate. Figure 10-7 illustrates such a control system.

10.3.2 Hydrated Lime

10.3.2.1 Properties and Availability

Hydrated lime, $Ca(OH)_2$, has a working density of 20 to 50 lb/ft^3, tends to flood the feeder, and will arch in storage bins if packed. The molecular weight is 74.08. The dust and slurry of hydrated lime are caustic in nature. The 1973 cost of bulk hydrated lime varies from $20 to $22 per ton. Bagged lime is available but increases the cost about $5 per ton. The availability of hydrated lime may be determined by contracting manufacturers listed in a subsequent section. The pH of a saturated hydrated lime solution is the same as that given for quicklime.

10.3.2.2 General Design Considerations

Hydrated lime is slaked lime and needs only enough water added to form milk of lime. Wetting or dissolving chambers are usually designed to provide 5 minutes detention with a ratio of 0.5 lb per gallon of water or 6% slurry at the maximum feed rate. Hydrated lime is usually used where maximum feed rates do not exceed 250 lb/hour. Hydrated lime and milk of lime will irritate the eyes, nose, and respiratory system and will dry the skin. Affected areas should be washed with water.

10.3.2.3 Storage

Information given for quicklime also applies to hydrated lime except that bin agitation must be provided. Bulk bin outlets should be provided with non-flooding rotary feeders and with hopper slopes of 65%.

10.3.2.4 Feeding Equipment

Volumetric or gravimetric feeders may be used, but volumetric feeders are usually selected only for installations where comparatively low feed rates are required. Dilution does not appear to be important; therefore, control of the amount of water used in the feeding operation is not considered necessary. Inexpensive hydraulic jet agitation may be furnished in the wetting chamber of the feeder as an alternative to mechanical agitation. The jets should be sized for the available water supply pressure to obtain proper mixing.

10.3.2.5 Piping and Accessories

Piping and accessories as described for quicklime are also appropriate for hydrated lime.

10.3.2.6 Pacing and Control

Controls as listed for dry alum apply to hydrated lime. Hydraulic jets should operate continuously and only shut off when the feeder is taken out of service. Control of the feed rate with pH as well as pacing with the plant flow may be used with hydrated lime as well as quicklime.

10.3.3 Estimated Initial Costs of Adding Lime Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of chemical storage and feeding equipment have been prepared based on the use of hydrated lime for a 1 mgd plant, and pebble quicklime for 10 and 100 mgd plants. Lime feed rates are based on a dosage of 150 mg/l and allow for peak rates of twice this capacity. Storage was provided for at least 15 days at the average rate although it may be desirable to provide 30 days for dry chemicals. This storage time is arbitrary and will vary at each installation depending on the distance to and the reliability of the source of chemical supply. Piping and buildings to house the feeding equipment are not included in the estimates. The estimated costs of steel bins with dust collector vents and filling accessories are included for all three plants. The chemical feeding equipment can be installed in existing buildings at many locations. In all estimates, the costs include an allowance for the contractor's installation, overhead, and profit, and an allowance of 20% of the construction cost of engineering and contingencies. The estimated costs for each size plant are shown in Table 10-8. These figures correspond to an EPA Construction Cost Index of 185.

TABLE 10-8

COST OF LIME FEED FACILITIES

Plant Size (mgd)	Estimated Base Bid ($)	Additional Cost of Flow and pH Controls ($)	Total Cost ($)
1	19,000	5,000	24,000
10	95,000	15,000	110,000
100	365,000	100,000	475,000

Equipment for the 1.0 mgd plant includes two volumetric feeder systems with manually loaded bins, dissolving chambers and accessories. Equipment for the 10.0 mgd plant includes two gravimetric feeder-slaker systems. Steel bins are estimated to hold 1-1/2 truckloads of quicklime each, and it is assumed that the truck will be equipped with a pneumatic blower. Equipment for the 100.00 mgd plant includes four gravimetric feeder-slaker systems. The base estimate includes the bin gate and feeder-slaker accessories and controls.

Costs for the flow and pH controls include holding tanks with mixers, and slurry feeders with controls. Flow and pH metering instruments are not included.

10.4 Compounds for pH Adjustment

10.4.1 Soda Ash

10.4.1.1 Properties and Availability

Soda ash, Na_2CO_3, is available in two forms. Light soda ash has a bulk density range of 35 to 50 lb/ft^3 and a working density of 41 lb/ft^3. Dense soda ash has a density range of 60 to 76 lb/ft^3 and a working density of 63 lb/ft^3. The pH of a 1% solution of soda ash is 11.2.

The molecular weight of soda ash is 106. Commercial purity ranges from 98 to 99%. The viscosities of sodium carbonate solutions are given in Figure 10-8. Soda ash by itself is not particularly corrosive, but in the presence of lime and water, caustic soda is formed which is quite corrosive.

Soda ash is available in bulk by truck, box car and hopper car, and in 100 lb bags from locations listed in a subsequent section. The 1974 price for soda ash ranges from $50 to $60 per ton, F.O.B. the point of manufacture; however, prices vary substantially between manufacturers and should be obtained from the closest manufacturers or local distributors.

10.4.1.2 General Design Considerations

Dense soda ash is generally used in municipal applications because of superior handling characteristics. It has little dust, good flow characteristics, and will not arch in the bin or flood the feeder. It is relatively hard to dissolve and ample dissolver capacity must be provided. Normal practice calls for 0.5 lb of dense soda ash per gallon of water or a 6% solution retained for 20 minutes in the dissolver.

The dust and solution are irritating to the eyes, nose, lungs and skin and therefore general precautions should be observed and the affected areas should be washed promptly with water.

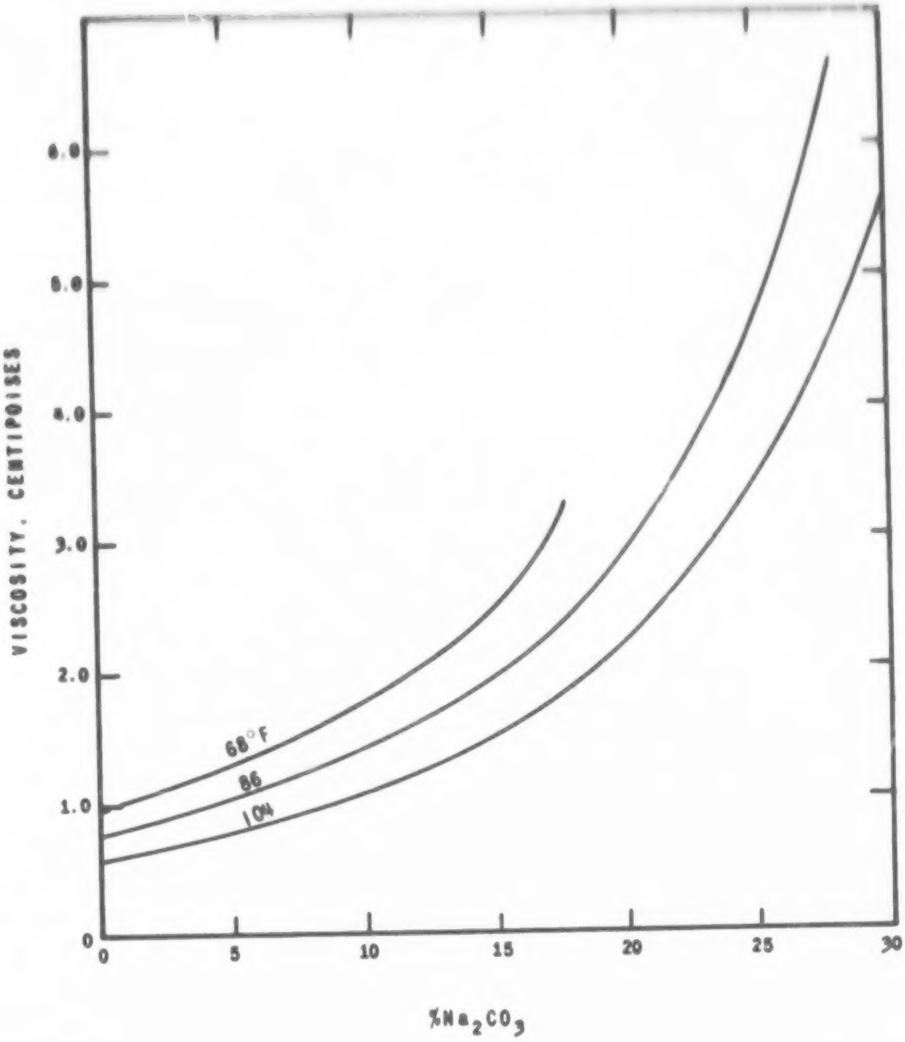
10.4.1.3 Storage

Soda ash is usually stored in steel bins and where pneumatic filling equipment is used, bins should be provided with dust collectors. Bulk and bagged soda ash tend to absorb atmospheric carbon dioxide and water forming the less active sodium bicarbonate ($NaHCO_3$). Material recommended for unloading facilities is steel.

FIGURE 10-8

VISCOSITY OF SODA ASH SOLUTIONS

(COURTESY PPG INDUSTRIES INC., CHEMICAL DIV.)



10.4.1.4 Feeding Equipment

Feed equipment as described for dry alum is suitable for soda ash. Dissolving of soda ash may be hastened by the use of warm dissolving water. Mechanical or hydraulic jet mixing should be provided in the dissolver.

10.4.1.5 Piping and Accessories

Materials of construction for piping and accessories should be iron, steel, rubber, and plastics.

10.4.1.6 Pacing and Control

Controls as discussed for dry alum apply also to soda ash equipment.

10.4.2 Liquid Caustic Soda

Anhydrous caustic soda (NaOH) is available but its use is generally not considered practical in water and wastewater treatment applications. Consequently, only liquid caustic soda is discussed below.

10.4.2.1 Properties and Availability

Liquid caustic soda is shipped at two concentrations, 50% and 73% NaOH. The densities of the solutions as shipped are 12.76 lb/gal for the 50% solution and 14.18 lb/gal for the 73% solution. These solutions contain 6.38 lb/gal NaOH and 10.34 lb/gal NaOH, respectively. The crystallization temperature is 53°F for the 50% solution and 165°F for the 73% solution. The molecular weight of NaOH is 40. Viscosities of various caustic soda solutions are presented in Figure 10-9. The pH of a 1% solution of caustic soda is 12.9.

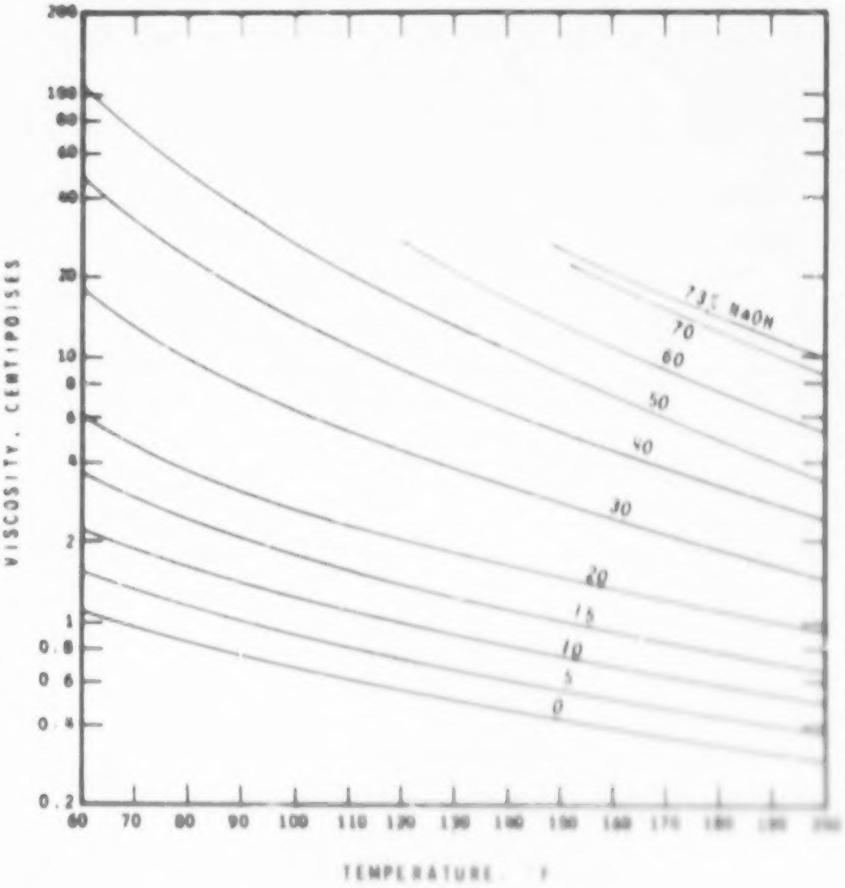
Truckload lots of 1,000 to 4,000 gallons are available in the 50% concentration only. Both shipping concentrations can be obtained in 8,000, 10,000, and 16,000 gallon carload lots. Tank cars can be unloaded through the dome eduction pipe using air pressure or through the bottom valve by gravity or by using air pressure or a pump. Trucks are usually unloaded by gravity or with air pressure or a truck mounted pump.

Major producers of caustic soda and their respective plant locations are listed in a subsequent section. The current price for liquid caustic soda ranges from about $75 to $80 per ton (NaOH), F.O.B. the point of manufacture.

FIGURE 10-9

VISCOSITY OF CAUSTIC SODA SOLUTIONS

(COURTESY OF HOOKER CHEMICAL CO.)



10.4.2.2 General Design Considerations

Liquid caustic soda is received in bulk shipments, transferred to storage and diluted as necessary for feeding to the points of application. Caustic soda is poisonous and is dangerous to handle. U.S. Department of Transportation Regulations for "White Label" materials must be observed. However, if handled properly caustic soda poses no particular industrial hazard. To avoid accidental spills, all pumps, valves, and lines should be checked regularly for leaks. Workmen should be thoroughly instructed in the precautions related to the handling of caustic soda. The eyes should be protected by goggles at all times when exposure to mist or splashing is possible. Other parts of the body should be protected as necessary to prevent alkali burns. Areas exposed to caustic soda should be washed with copious amounts of water for 15 minutes to 2 hours. A physician should be called when exposure is severe. Caustic soda taken internally should be diluted with water or milk and then neutralized with dilute vinegar or fruit juice. Vomiting may occur spontaneously but should not be induced except on the advice of a physician.

10.4.2.3 Storage

Liquid caustic soda may be stored at the 50% concentration. However, at this solution strength, it crystallizes at 53°F. Therefore, storage tanks must be located indoors or provided with heating and suitable insulation if outdoors. Because of its relatively high crystallization temperature, liquid caustic soda is often diluted to a concentration of about 20% NaOH for storage. A 20% solution of NaOH has a crystallization temperature of about -20°F. Recommendations for dilution of both 73% and 50% solutions should be obtained from the manufacturer, because special considerations are necessary.

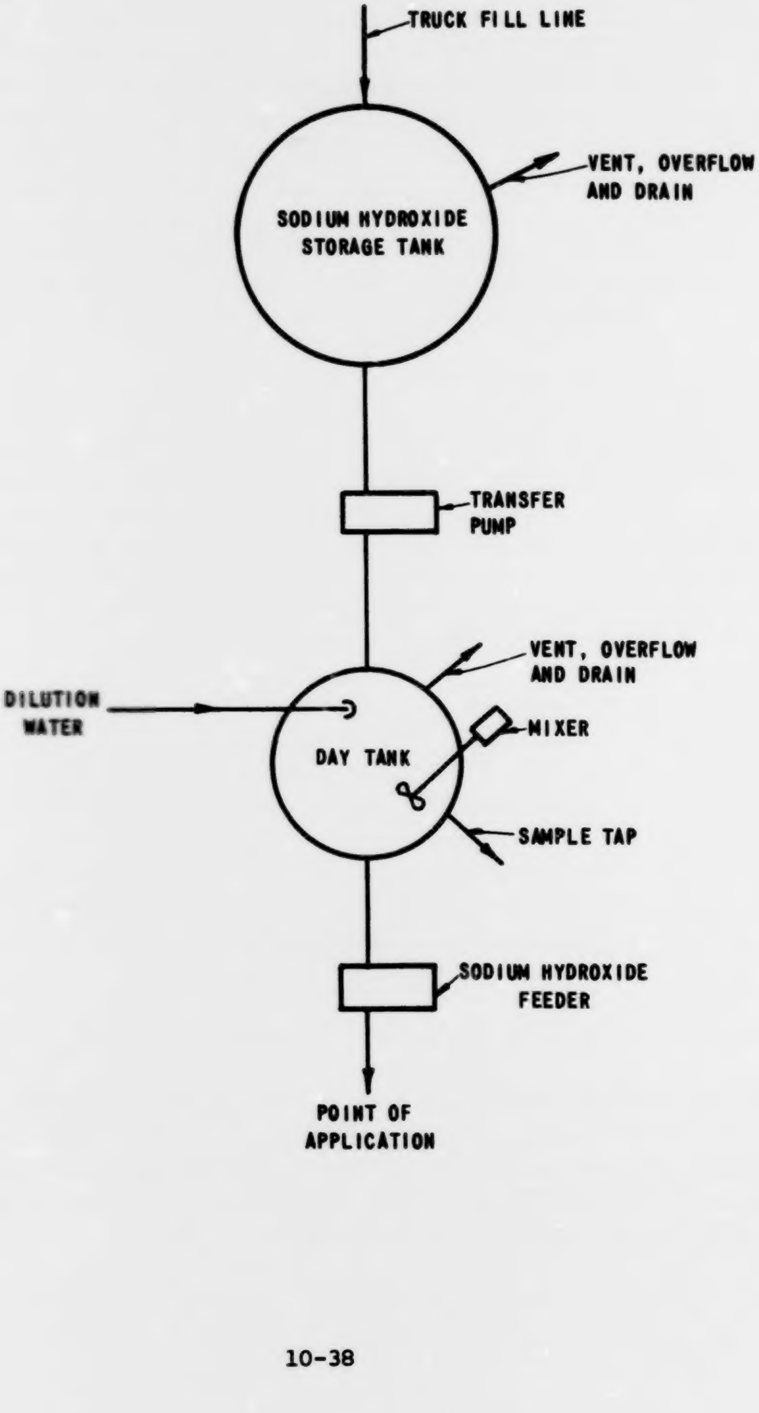
Storage tanks for liquid caustic soda should be provided with an air vent for gravity flow. The storage capacity should be equal to 1-1/2 times the largest expected delivery, with an allowance for dilution water, if used, or 2 weeks supply at the anticipated feed rate, whichever is greater. Tanks for storing 50% solution at a temperature between 75°F and 140°F may be constructed of mild steel. Storage temperatures above 140°F require more elaborate materials selection and are not recommended. Caustic soda will tend to pick up iron when stored in steel vessels for extended periods. Subject to temperature and solution strength limitations, rubber, 316 stainless steel, nickel, nickel alloys, or plastics may be used when iron contamination must be avoided.

10.4.2.4 Feeding Equipment

Further dilution of liquid caustic soda below the storage strength may be desirable for feeding by volumetric feeders. Feeding systems as described for liquid alum generally apply to caustic soda with appropriate selection of materials of construction. A typical system schematic is shown in Figure 10-10. Feeders will usually include materials such as ductile iron, stainless steels, rubber, and plastics.

FIGURE 10-10

TYPICAL CAUSTIC SODA FEED SYSTEM


TRUCK FILL LINE
VENT, OVERFLOW AND DRAIN
SODIUM HYDROXIDE STORAGE TANK
TRANSFER PUMP
VENT, OVERFLOW AND DRAIN
DILUTION WATER
DAY TANK
MIXER
SAMPLE TAP
SODIUM HYDROXIDE FEEDER
POINT OF APPLICATION

10.4.2.5 Piping and Accessories

Transfer lines from the shipping unit to the storage tank should be spiral-wire-bound neoprene or rubber hose, solid steel pipe with swivel joints, or steel hose. Because caustic soda attacks glass, use of glass materials should be avoided. Other miscellaneous materials for use with liquid caustic soda feeding and handling equipment are listed below (7) in Table 10-9:

TABLE 10-9

MATERIALS SUITABLE FOR CAUSTIC SODA SERVICE

Components	Recommended Materials for Use With 50% NaOH Up to 140°F
Rigid Pipe	Standard Weight Black Iron
Flexible Connections	Rigid Pipe with Ells or Swing Joints, Stainless Steel or Rubber Hose
Diluting Tees	Type 304 Stainless Steel
Fittings	Steel
Permanent Joints	Welded or Screwed Fittings
Unions	Screwed Steel
Valves - Non-leaking (Plug)	
Body	Steel
Plug	Type 304 Stainless Steel
Pumps (Centrifugal)	
Body	Steel
Impeller	Ni-Resist
Packing	Blue Asbestos
Storage Tanks	Steel

10.4.2.6 Pacing and Control

Controls as listed for liquid alum also apply to liquid caustic soda equipment.

10.5 Carbon Dioxide

10.5.1 Properties and Availability

Carbon dioxide, CO_2, is available for use in wastewater treatment plants in gas and liquid form. The molecular weight of carbon dioxide is 44. Dry carbon dioxide is not chemically active at normal temperatures and is a non-toxic safe chemical; however, the gas displaces oxygen and adequate ventilation of closed areas should be provided. Solutions of carbon dioxide in water are very reactive chemically and form carbonic acid. Saturated solutions of carbon dioxide have a pH of 4.0 at 68°F.

The gas form may be produced on the treatment plant site by scrubbing and compressing the combustion product of lime recalcining furnaces, sludge furnaces, or generators used principally for the production of carbon dioxide gas only. These generators are usually fired with combustible gases, fuel oil, or coke and have carbon dioxide yields as shown in Table 10-10 (8).

TABLE 10-10

CARBON DIOXIDE YIELDS OF COMMON FUELS

Fuel	Quantity	Yield lb
Natural Gas	1,000 ft^3	115
Coke	1 lb	3
Kerosene	1 gal	20
Fuel Oil (No. 2)	1 gal	23
Propane	1,000 ft^3	352
Butane	1,000 ft^3	454

The gas forms, as generated at the plant site, usually have a carbon dioxide content of between 6% and 18% depending on the source and efficiency of the producing system.

The liquid form is available from commercial suppliers in 20 to 50 lb cylinders, 10 to 20 ton trucks and 30 to 50 ton rail cars. The commercial liquid form has a minimum carbon dioxide content of 99.5%.

1974 prices range from $33-$36 per ton for 3,000 tons per year and over, to $71-$74 per ton for a quantity of 150 tons per year. These prices include an allowance for freight within a 100 mile radius of the point of manufacture. Another $6 or $7 per ton may be added for each additional 100 miles to the point of destination. Major producers of commercial carbon dioxide are listed in a subsequent section.

10.5.2 General Design Considerations

Recovery of carbon dioxide from recalcining furnaces or incinerators is the least expensive source, but maintenance of stack gas systems is likely to be extensive because of the corrosive nature of the wet gas and the presence of particulate matter. Scrubber systems are required to clean the stack gas and especially designed gas compressors are necessary to provide the process injection pressure.

Pressure generators and submerged burners require less maintenance because the system pressure is established by compressors or blowers handling dry air or gas. On-site generating units have a limited range of carbon dioxide production as compared with the liquid storage and feed system, and therefore may require multiple units.

The liquid carbon dioxide storage and feed system generally includes a temperature-pressure controlled, bulk storage tank, an evaporation unit, and a gas feeder to meter the gas. Solution feeders, similar in construction to chlorinators, may also be used to feed carbon dioxide.

10.5.3 Storage

This section applies only to use of commercial liquid carbon dioxide. Liquid system capacities encountered in wastewater treatment usually usually require on-site bulk storage units. Standard pre-packaged units are available, ranging in size from 3/8 to 50 tons capacity, and are furnished with temperature-pressure controls to maintain approximately 300 lb/in^2 at 0^oF conditions. The typical package unit contains refrigeration, vaporization, safety and control equipment. The units are well insulated and protected for outdoor location. The gas from the evaporation unit usually passes through two stages of pressure reduction before entering the gas feeder to prevent the formation of dry ice.

10.5.4 Feeding Equipment

Feeding systems for the stack gas source of carbon dioxide consist of simple valving arrangements, for admitting varying quantities of make-up air to the suction side of the constant volume compressors, or for venting excess gas on the compressor discharge. A typical system is described by Culp and Culp (9).

Pressure generators and submerged burners are regulated by valving arrangements on the fuel and air supply. Generation of carbon dioxide by combustion is usually difficult to control, requires frequent operator attention and demands considerable maintenance over the life of the equipment, when compared to liquid CO_2 systems.

Commercial liquid carbon dioxide is becoming more generally used because of its high purity, the simplicity and range of feeding equipment, ease of control, and smaller, less expensive piping systems. After vaporization, carbon dioxide with suitable metering and pressure reduction may be fed directly to the point of application as a gas. Metering of direct fed pressurized gas is difficult due to high adiabatic expansion characteristics of the gas. Also, direct feed requires extremely fine bubbles to insure the gas goes into solution; this, in turn, can lead to scaling problems. However, vacuum operated, solution type gas feeders are preferred. Such feeders generally include safety devices and operating controls in a compact panel housing, with materials of construction suitable for carbon dioxide service. Absorption of carbon dioxide in the injector water supply approaches 100% when a ratio of 1.0 lb of gas to 60 gal of water is maintained.

10.5.5 Piping and Accessories

Mild steel piping and accessories are suitable for use with cool, dry, carbon dioxide. Hot, moist gases, however, require the use of type 316 stainless steel or plastic materials. Plastics or FRP pipe are generally used for solution piping and diffusers. Diffusers should be submerged at least 8 ft, and preferably deeper, to assure complete absorption of the gas.

10.5.6 Pacing and Control

Standard instrument signals and control components can be used to pace or control carbon dioxide feed systems.

Using stack gas as the source of carbon dioxide, the feed rate can be controlled by proper selection and operation of compressors, by manual control of vent or bleed valves, or by automatic control of such valves by a pH meter-controller system.

In commercial carbon dioxide feed systems, solution feeders may function as controllers and can be paced by instrument signals from pH monitors and plant flow meters.

In feeding commercial carbon dioxide directly to the point of application as a gas, a differential pressure transmitter and a control valve may function as the primary elements of a control system. Standard instrument signals may be used to pace or control the rate of carbon dioxide feed.

Carbon dioxide generators are difficult to pace or control other than by manual or automatic operation of vent or bleed valves that waste a portion of the produced gas according to the plant requirements.

10.6 Polymers

10.6.1 Dry Polymers

10.6.1.1 Properties and Availability

Types of polymers vary widely in characteristics. Manufacturers should be consulted for properties, availability, and cost of the polymer being considered. References are available that indicate the types and characteristics of polymers available (10, 11). Bulk shipments are generally not desirable. Polymers are available in a variety of container or package sizes.

10.6.1.2 General Design Considerations

Dry polymer and water must be blended and mixed to obtain a recommended solution for efficient action. Solution concentrations vary from fractions of a percent up. Preparation of the stock solution involves wetting of the dry material and usually an aging period prior to application. Solutions can be very viscous, and close attention should be paid to piping size and length and pump selections. Metered solution is usually diluted just prior to injection to the process to obtain better dispersion at the point of application.

10.6.1.3 Storage

General practice for storage of bagged dry chemicals should be observed. The bags should be stored in a dry, cool, low humidity area and used in proper rotation, i.e., first in, first out.

Solutions are generally stored in type 316 stainless steel, FRP, or plastic lined tanks.

10.6.1.4 Feeding Equipment

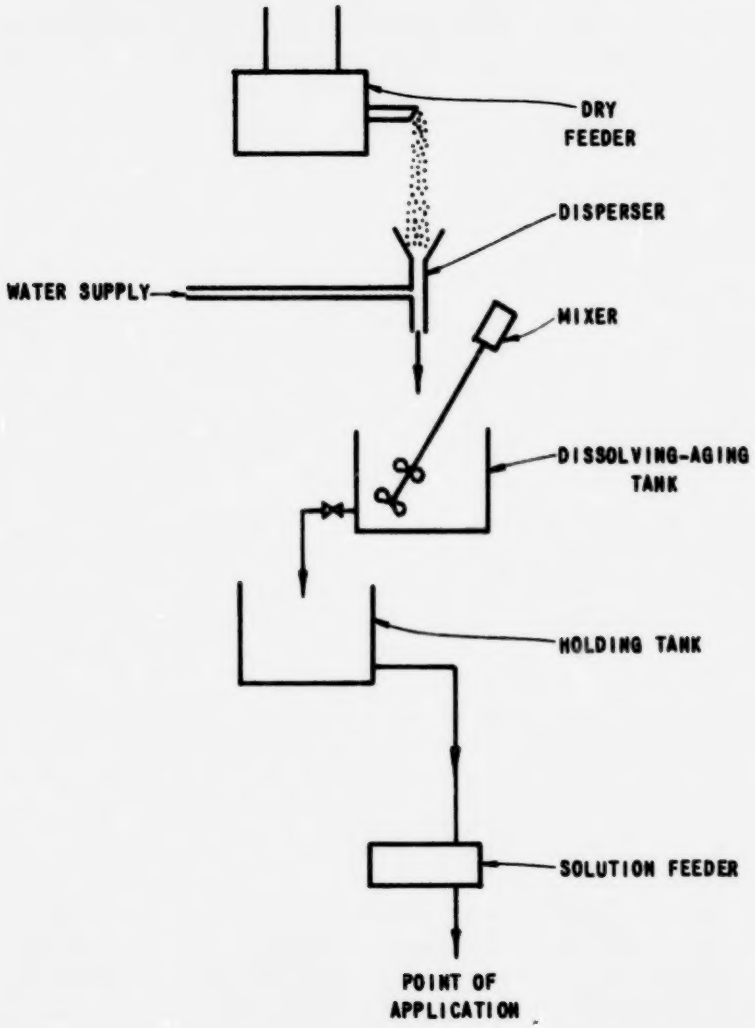
Two types of systems are frequently combined to feed polymers. The solution preparation system includes a manual or automatic blending system with the polymer dispensed by hand or by a dry feeder to a wetting jet and then to a mixing-aging tank at a controlled ratio. The aged polymer is transported to a holding tank where metering pumps or rotodip feeders dispense the polymer to the process. A schematic of such a system is shown by Figure 10-11. It is generally advisable to keep the holding or storage time of polymer solutions to a minimum, 1 to 3 days or less, to prevent deterioration of the product.

10.6.1.5 Piping and Accessories

Selection must be made after determination of the polymer; however, type 316 stainless steel or plastics are generally used.

FIGURE 10-11

MANUAL DRY POLYMER FEED SYSTEM


DRY FEEDER
DISPERSER
WATER SUPPLY
MIXER
DISSOLVING-AGING TANK
HOLDING TANK
SOLUTION FEEDER
POINT OF APPLICATION

10.6.1.6 Pacing and Control

Controls as listed for liquid alum apply to the control of liquid dispersing feeders.

The solution preparation system may be an automatic batching system, as shown by the schematic on Figure 10-12, that fills the holding tank with aged polymer as required by level probes. Such a system is usually provided only at large plants. Unitized solution preparation units are available, but have a limited capacity.

10.6.2 Liquid Polymers

10.6.2.1 Properties and Availability

As with dry polymers, there is a wide variety of products, and manufacturers should be consulted for specific information.

10.6.2.2 General Design Considerations

Liquid systems differ from the dry systems only in the equipment to blend the polymer with water to prepare the solution. Liquid solution preparation is usually a hand-batching operation with manual filling of a mixing-aging tank with water and polymer.

10.6.2.3 Feeding Equipment

Liquid polymers need no aging and simple dilution is the only requirement for feeding. The dosage of liquid polymers may be accurately controlled by metering pumps or rotodip feeders.

The balance of the process is generally the same as described for dry polymers.

10.6.3 Estimated Initial Costs of Adding Polymer Feed Facilities to Existing Wastewater Treatment Plants

Cost estimates of solution preparation and feeder equipment have been prepared, based on the use of dry polymer, for 1, 10, and 100 mgd plants. Chemical feed equipment was sized to have a capacity sufficient to prepare and feed a 0.25% stock solution at a dosage of 1 mg/l. Piping and buildings to house the feeding equipment and store the bags are not included. At many locations, the chemical feeding equipment can be installed in existing buildings. All estimated costs include an allowance for the contractor's installation, overhead, and profit plus an allowance of 20% of the construction cost for engineering and contingencies. The estimated costs for each size plant are shown in Table 10-11. These figures correspond to an EPA Construction Cost Index of 185:

FIGURE 10-12

AUTOMATIC DRY POLYMER FEED SYSTEM

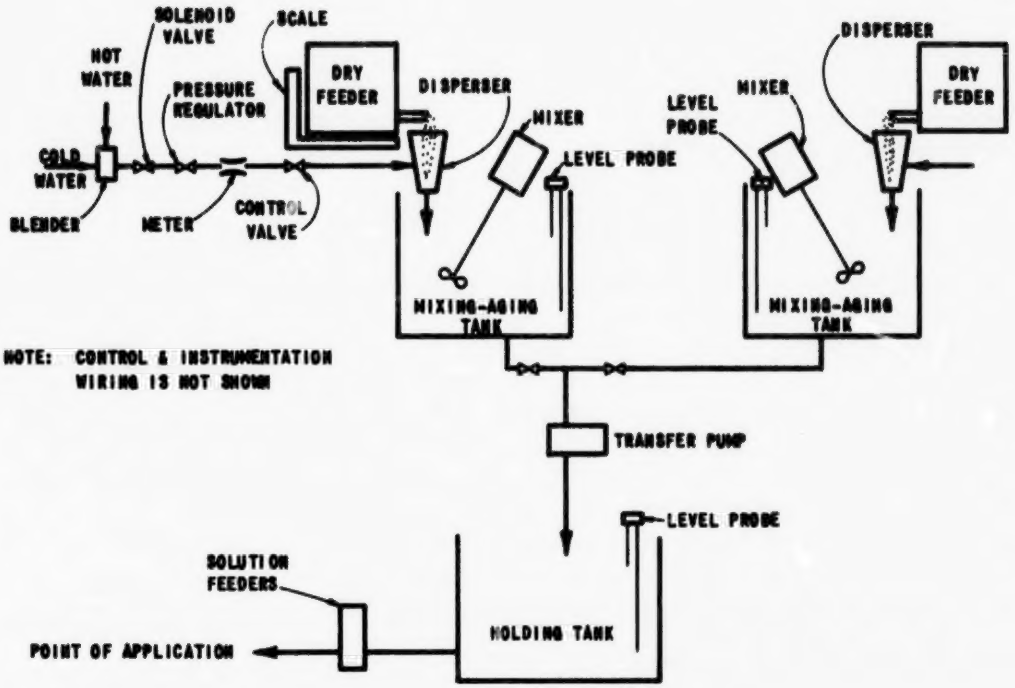

SOLENOID VALVE
SCALE
HOT WATER
DRY FEEDER
PRESSURE REGULATOR
DISPERSER
MIXER
LEVEL PROBE
COLD WATER
BLENDER
METER
CONTROL VALVE
MIXING-AGING TANK
LEVEL PROBE
MIXER
DISPERSER
DRY FEEDER
MIXING-AGING TANK
NOTE: CONTROL & INSTRUMENTATION WIRING IS NOT SHOWN
TRANSFER PUMP
LEVEL PROBE
SOLUTION FEEDERS
HOLDING TANK
POINT OF APPLICATION

TABLE 10-11

COST OF POLYMER FEED FACILITIES

Plant Size (mgd)	Estimated Cost ($)
1	7,000
10	31,000
100	225,000

The system for the 1 mgd plant was based on the polymer being manually fed to the mixing tank, with mixing and solution feed equipment arranged for manual control. Two independent systems of tanks and feeders are included. The system for the 10 mgd plant includes two volumetric dry feeders discharging to two mixing tanks, arranged for batch control. The mixing tanks discharge to a single holding tank from which two solution feeders take their supply for application to the process. The system for the 100 mgd plant includes four volumetric dry feeders complete with steel day hoppers, dust collectors, bin gates and flexible connectors. The system operation is automatic for the preparation and transfer of aged polymer solution from four mixing tanks to two holding tanks. Ten solution feeders meter the polymer to the treatment process.

10.7 Chemical Suppliers

10.7.1 Metal Salts and Lime

The first and largest list of suppliers concerns metal salts and lime. It is divided geographically into states, which are tabulated alphabetically. Then, under each state, production locations are listed for the following alphabetized chemicals:

alum (dry and/or liquid)

ferric chloride

ferric sulfate

ferrous sulfate

lime (high calcium and/or dolomitic)

sodium aluminate (dry and/or liquid)

Table 10-12 summarizes extent and location of presently available sources of metal salts and lime:

TABLE 10-12

GEOGRAPHICAL SUMMARY OF SUPPLIERS OF METAL SALTS AND LIME

Location	Chemical
25 States	Alum
Michigan	Ferric chloride
Tennessee	Ferric sulfate
Ga., Md., Mo., N.J.	Ferrous sulfate
32 states	Lime
Ark., Ill., La., Ga., Mo.	Sodium aluminate

Table 10-13 offers a detailed listing of metal salt and lime suppliers. Persons interested in contacting chemical suppliers may direct inquiries to either the point of production or to a nearby sales office for the chemical company involved. This list does not include potential or established suppliers of waste pickle liquor. Nor are pH adjusting chemicals included here; these are normally available through most major chemical supply firms.

TABLE 10-13

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
ALABAMA		
Coosa Pines	American Cyanamid	Alum, Liquid
Demopolis	American Cyanamid	Alum, Liquid
Mobile	American Cyanamid	Alum, Liquid and Dry
Naheola	Stauffer	Alum, Liquid
Allgood	Cheney Lime & Cement Co.	Lime, High Calcium
Keystone	Martin Marietta Cement, Southern Division	Lime, High Calcium
Landmark	Cheney Lime & Cement Co.	Lime, High Calcium
Montevallo	U.S. Gypsum Co.	Lime, High Calcium
Roberta	Southern Cement Co., Div. Martin Marietta Corp.	Lime, High Calcium
Saginaw	Longview Lime Co., Div. Woodward Co., Div. Mead Corp.	Lime, High Calcium
Siluria	Alabaster Lime Co.	Lime, High Calcium
ARIZONA		
Douglas	Paul Lime Plant, Inc.	Lime, High Calcium
Globe	Hoopes & Co.	Lime, High Calcium
Nelson	U.S. Lime Div., The Flintkote Co.	Lime, High Calcium
ARKANSAS		
Pine Bluff	Allied	Alum, Liquid
Batesville	Batesville White Lime Co., Div. Rangaire Corp.	Lime, High Calcium
Bauxite	Reynolds Chemicals	Sodium Aluminate, Dry

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
CALIFORNIA		
Bay Point	Allied	Alum, Liquid and Dry
Antioch	Imperial West Chemicals	Ferric Chloride
El Segundo	Allied	Alum, Liquid
Richmond	Stauffer	Alum, Liquid
Vernon	Stauffer	Alum, Liquid
City of Industry	U.S. Lime Div., The Flintkote Co.	Lime, High Calcium
Diamond Springs	Diamond Springs Lime Co.	Lime, High Calcium
Lucerne Valley	Pfizer, Inc., Minerals, Pigments & Metals Div.	Lime, High Calcium
Richmond	U.S. Lime Div., The Flintkote Co.	Lime, High Calcium
Salinas	Kaiser Aluminum & Chemical Corp.	Lime, Dolomitic
Westend	Stauffer Chemical Co.	Lime, High Calcium
COLORDAO		
Denver	Allied	Alum, Liquid and Dry
Ft. Morgan	Great Western Sugar Co.	Lime, High Calcium
CONNECTICUT		
Canaan	Pfizer, Inc., Minerals Pigments & Metals Div.	Lime, Dolomitic
FLORIDA		
Fernandina Beach	Tennessee Corp.	Alum, Liquid
Jacksonville	Allied	Alum, Liquid
Port St. Joe	Allied	Alum, Liquid

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
FLORIDA (cont.)		
Brooksville	Chemical Lime Co.	Lime, High Calcium
Sumterville	Dixie Lime and Stone Co.	Lime, High Calcium
GEORGIA		
Atlanta (2)	Burris; Allied	Alum, Liquid and Dry
Augusta	Tennessee Corp.	Alum, Liquid and Dry
Cedar Springs	Tennessee Corp.	Alum, Liquid
Macon	Allied	Alum, Liquid
Savannah	Allied	Alum, Liquid
Savannah	American Cyanamid Co.	Ferrous Sulfate
Vinings	Vinings Chemical Co.	Sodium Aluminate, Liquid
ILLINOIS		
E. St. Louis	Allied	Alum, Liquid and Dry
Joliet	American Cyanamid	Alum, Liquid and Dry
Marblehead	Marblehead Lime Co.	Lime, High Calcium
McCook	Vulcan Materials Co.	Lime, Dolomitic
Quincy	Marblehead Lime Co.	Lime, High Calcium
So. Chicago	Marblehead Lime Co.	Lime, High Calcium
Thornton	Marblehead Lime Co.	Lime, Dolomitic
Chicago	Nalco Chemical Company	Sodium Aluminum, Liquid and Dry
INDIANA		
Buffington	Marblehead Lime Co.	Lime, High Calcium

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
IOWA		
Davenport	Linwood Stone Products Co., Inc.	Lime, High Calcium
KENTUCKY		
Carntown	Black River Mining Co.	Lime, High Calcium
LOUISIANA		
Bastrop	Stauffer	Alum, Liquid
Baton Rouge	Stauffer	Alum, Liquid
Monroe	Allied	Alum, Liquid
New Orleans	Allied	Alum, Liquid and Dry
Springhill	Stauffer	Alum, Liquid
Morgan City	Pelican State Lime Corp.	Lime, High Calcium
New Orleans	U.S. Gypsum Co.	Lime, High Calcium
MAINE		
Searsport	Northern	Alum, Liquid and Dry
MARYLAND		
Baltimore	Olin	Alum, Dry
Baltimore (3)	Byproducts Processing Co., Inc.;	Ferrous Sulfate
	Glidden Co.;	Ferrous Sulfate
	Cosmin Corp.;	Ferrous Sulfate
Woodsboro	S.W. Barrick & Sons, Inc.	Lime, High Calcium

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
MASSACHUSETTS		
Adams	Holland	Alum, Liquid
Salem	Hamblet & Hayes	Alum, Liquid and Dry
Adams	Pfizer, Inc., Minerals Pigments & Metals Div.	Lime, High Calcium
Lee	Lee Lime Corp.	Lime, Dolomitic
MICHIGAN		
Detroit	Allied	Alum, Liquid
Escanaba	American Cyanamid	Alum, Liquid
Kalamazoo (2)	Allied; American Cyanamid	Alum, Liquid
Midland	Dow Chemical Co.	Ferric Chloride
Wyandotte	Pennwalt Corp.	Ferric Chloride
Detroit	Detroit Lime Co.	Lime, High Calcium
Ludington	Dow Chemical Co.	Lime, High Calcium
River Rouge	Marblehead Lime Co.	Lime, High Calcium
Cloquet	American Cyanamid	Alum, Liquid
Pine Bend	North Star	Alum, Liquid and Dry
Duluth	Cutler Magner Co.	Lime, High Calcium
MISSISSIPPI		
Monticello	American Cyanamid	Alum, Liquid
Vicksburg	Allied	Alum, Liquid

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
MISSOURI		
St. Louis	National Lead	Ferrous Sulfate
Bonne Terre	Valley Mineral Products Co.	Lime, Dolomitic
Hannibal	Marblehead Lime Co.	Lime, High Calcium
Ste. Genevieve	Mississippi Lime Co.	Lime, High Calcium
Springfield	Ash Grove Cement Co.	Lime, High Calcium
Kansas City	Conservation Chemical Co.	Sodium Aluminate, Liquid
NEVADA		
Apex	U.S. Lime Div., The Flintkote Co.	Lime, High Calcium
Henderson	U.S. Lime Div., The Flintkote Co.	Lime, Dolomitic and High Calcium
McGill	Morrison-Weatherly Corp.	Lime, High Calcium
Sloan	U.S. Lime Div., The Flintkote Co.	Lime, Dolomitic and High Calcium
NEW JERSEY		
Newark	Essex	Alum, Liquid and Dry
Warners	American Cyanamid	Alum, Liquid and Dry
Sayreville	National Lead	Ferrous Sulfate
Newton	Limestone Products Corp. of America	Lime, High Calcium, Hydrate Only

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
NORTH CAROLINA		
Acme	Wright	Alum, Liquid
Plymouth	American Cyanamid	Alum, Liquid
OHIO		
Chillicothe	Allied	Alum, Liquid
Cleveland	Allied	Alum, Liquid and Dry
Hamilton	American Cyanamid	Alum, Liquid and Dry
Middletown	Allied	Alum, Liquid
Ashtabula	Union Carbide Olefins Co.	Lime, High Calcium
Carey	National Lime & Stone Co.	Lime, Dolomitic
Cleveland	Cuyahoga Lime Co.	Lime, High Calcium
Delaware	MCQ Co.	Lime, High Calcium
Genoa	U.S. Gypsum Co.	Lime, Dolomitic
Gibsonburg (2)	Pfizer, Inc., Minerals, Pigments & Metal Div.; National Gypsum Co.	Lime, Dolomitic
Huron	Huron Lime Co.	Lime, High Calcium
Marble Cliff	MCQ Co.	Lime, High Calcium
Millersville	J.E. Baker Co.	Lime, Dolomitic
Woodville (2)	Ohio Lime Co.; Standard Lime & Refractories Div., Martin Marietta Corp.	Lime, Dolomitic
OKLAHOMA		
Marble City	St. Clair Lime Co.	Lime, High Calcium
Sallisaw	St. Clair Lime Co.	Lime, High Calcium

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
OREGON		
North Portland	Stauffer	Alum, Liquid and Dry
Baker	Chemical Lime Co. of Oregon	Lime, High Calcium
Portland	Ash Grove Cement Co.	Lime, High Calcium
PENNSYLVANIA		
Johnsonburg	Allied	Alum, Liquid
Marcus Hook	Allied	Alum, Liquid and Dry
Newell	Allied	Alum, Liquid
Annville	Bethlehem Mines Corp.	Lime, High Calcium
Bellefonte (2)	National Gypsum Co.; Warner Co.	Lime, High Calcium
Branchton	Mercer Lime & Stone Co.	Lime, High Calcium
Devault	Warner Co.	Lime, Dolomitic
Pleasant Gap	Chemical Lime Inc.	Lime, High Calcium
Plymouth Meeting	G.&W.H. Corson, Inc.	Lime, Dolomitic
SOUTH CAROLINA		
Catawba	Burris	Alum, Liquid
Georgetown	American Cyanamid	Alum, Liquid and Dry
SOUTH DAKOTA		
Rapid City	Pete Lien & Sons, Inc.	Lime, High Calcium

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
TENNESSEE		
Chattanooga	American Cyanamid	Alum, Liquid and Dry
Counce	Stauffer	Alum, Liquid
Springfield	Burris	Alum, Liquid
Copper Hill	Cities Service Co.	Ferric Sulfate
Knoxville (2)	Foote Mineral Co.; Williams Lime Mfg. Co.	Lime, High Calcium
TEXAS		
Houston (2)	Stauffer; Ethyl	Alum, Liquid and Dry
Texas City	Gulf Chemical & Metallurgical	Ferrous Sulfate, Ferric Sulfate
Blum	Round Rock Lime Companies	Lime, High Calcium
Cleburne	Texas Lime Co., Div. Rangaire Corp.	Lime, High Calcium
Clifton	Chemical Lime Inc.	Lime, High Calcium
Houston	U.S. Gypsum Co.	Lime, High Calcium
McNeil	Austin White Lime Co.	Lime, High Calcium
New Braunfels	U.S. Gypsum Co.	Lime, High Calcium
Round Rock	Round Rock Lime Companies	Lime, High Calcium
San Antonio	McDonough Bros., Inc.	Lime, High Calcium
UTAH		
Grantsville	U.S. Lime Div., The Flintkote Co.	Lime, Dolomitic and High Calcium
Lehi	Rollins Mining Supplies Co.	Lime, High Calcium

SUPPLIERS OF METAL SALTS AND LIME

Location	Manufacturer	Chemical
VERMONT		
Winooski	Vermont Assoc. Lime Industries, Inc.	Lime, High Calcium
VIRGINIA		
Covington	Allied	Alum, Liquid
Hopewell	Allied	Alum, Liquid
Norfolk	Howerton Gowen	Alum, Liquid
Clearbrook	W.S. Frey Co., Inc.	Lime, High Calcium
Kimballton (2)	Foote Mineral Co.; National Gypsum Co.	Lime, High Calcium
Stephens City	M.J. Grove Lime Co., Div. The Flintkote Co.	Lime, High Calcium
Strasburg	Chemstone Corp.	Lime, High Calcium
WASHINGTON		
Kennewick	Allied	Alum, Liquid
Tacoma (2)	Allied; Stauffer	Alum, Liquid
Vancouver	Allied	Alum, Liquid and Dry
Tacoma	Domtar Chemicals Inc.	Lime, High Calcium
WEST VIRGINIA		
Riverton	Germany Valley Limestone Div., Greer Steel Co.	Lime, High Calcium

Location	Manufacturer	Chemical
WISCONSIN		
Manasha	Allied	Alum, Liquid
Wisconsin Rapids	Allied	Alum, Liquid
Eden	Western Lime & Cement Co.	Lime, Dolomitic
Green Bay	Western Lime & Cement Co.	Lime, High Calcium
Knowles	Western Lime & Cement Co.	Lime, Dolomitic
Manitowoc	Rockwell Lime Co.	Lime, Dolomitic
Superior	Cutler-LaLiberte-McDougall Corp.	Lime, High Calcium

10.7.2 Polymer Suppliers

Table 10-14 lists names and addresses of firms offering polymer which may prove effective in treatment approaches described herein.

TABLE 10-14

POLYMER SOURCES AND TRADE NAMES

Source	Trade Name(s)
Allied Colloids, Inc. One Robinson Lane Ridgewood, NJ 07450	Percol
Allstate Chemical Co. Post Office Box 3040 Euclid, OH 44117	Allstate

POLYMER SOURCES AND TRADE NAMES

Sources	Trade Name(s)
Allyn Chemical Co. 2224 Fairhill Road Cleveland, OH 44106	Claron
American Cyanimid Co. Berdan Avenue Wayne, NJ 07470	Superfloc Magnifloc
Atlad Chemical Ind., Inc. Wilmington, DE 19899	Sarbo Atlasep
Berdell Industries 28-01 Thomson Avenue Long Island City, NY 11101	Berdell
Betz Laboratories, Inc. Somerton Road Trevose, PA 19047	Betz Polyfloc
Bond Chemicals, Inc. 1500 Brookpark Road Cleveland, OH 44109	Bondfloc
Brennan Chemical Co. 704 N. First Street St. Louis, MO 63102	Brenco
The Burtonite Company Nutley, NJ 07110	Burtonite
Calgon Corporation Post Office Box 1346 Pittsburgh, PA 15222	Cat-Floc

POLYMER SOURCES AND TRADE NAMES

Source	Trade Name(s)
Commercial Chemical 11 Patterson Avenue Midland Park, NJ 07432	Speedifloc
Dearborn Chemical Div. W. R. Grace & Co. Merchandise Mart Plaza Chicago, IL 60654	Aquafloc
Dow Chemical USA Barstow Building 2020 Dow Center Midland, MI 48640	Dowell PEI Purifloc Separan XD
Drew Chemical Corp. 701 Jefferson Road Parsippany, NJ 07054	Drewfloc Amerfloc
DuBois Chemicals Div. W. R. Grace & Co. 3630 E. Kemper Road Sharonville, OH 45241	Flocculite
E. I. DuPont de Nemours & Co. Eastern Laboratory Gibbstown, NJ 08027	DuPont
Environmental Pollution Investigation & Control, Inc. 9221 Bond Street Overland Park, KS 66214	Dynafloc

POLYMER SOURCES AND TRADE NAMES

Source	Trade Name(s)
Fabcon International 1275 Columbus Avenue San Francisco, CA 94133	Zulcar Fabcon
Henry W. Fink & Co. 6900 Silverton Avenue Cincinnati, OH 45236	Kleer-Floc
Gamlen Sybron Corp. 321 Victory Avenue S. San Francisco, CA 94080	Gamafloc Gamlose Gamlen
Garrett-Callahan 111 Rollins Road Millbrae, CA	Garrett-Callahan
General Mills Chemicals 4620 N. 77th Street Minneapolis, MN 55435	Supercol Guartec
Hercules, Inc. 910 Market Street Wilmington, DE 19899	Hercofloc
Frank Herzl Corp. 299 Madison Avenue New York, NY 10017	Perfectamyl
ICI America, Inc. Wilmington, DE 19899	Atlasep
Illinois Water Treatment Co. 840 Cedar Street Rockford, IL 61102	Illco

POLYMER SOURCES AND TRADE NAMES

Source	Trade Name(s)
Kelco Company 8225 Aero Drive San Diego, CA 92123	Kelgin Kelcosol
Key Chemicals 4346 Tacony Philadelphia, PA 19124	Key-Floc
Metalene Chemical Co. Bedford, OH 44014	Metalene
The Mogul Corporation 20600 Chagrin Boulevard Cleveland, OH 44122	Mogul
Nalco Chemical Co. 6216 W. 66th Pl. Chicago, IL 60638	Nalcolyte
Narvon Mining & Chemical Co. Keller Avenue & Fruitville Pike Lancaster, PA 17604	Sink-Floc Zeta-Floc
National Starch & Chemical Corp. 1700 W. Front Street Plainfield, NJ 07063	Floc-Aid Natron
O'Brein Industries, Inc. 95 Dorsa Avenue Livingston, NJ 07039	O'B Floc
Oxford Chemical Div. Consolidated Foods Corp. Post Office Box 80202 Atlanta, GA 30341	Oxford-Hydro-Floc

Source	Trade Name(s)
Reichhold Chemicals, Inc. RCI Building White Plains, NY 10602	Aquarid
Standard Brands Chem. Ind., Inc. Post Office Drawer K Dover, DE 19901	Tychem
A. E. Staley Mfg. Co. Post Office Box 151 Decatur, IL 62525	Hamaco
Stein, Hall & Co., Inc. 605 Third Avenue New York, NY 10016	Hallmark Jaguar Polyhall
Swift & Company Oakbrook, IL 60521	Swift
James Varley & Sons, Inc. 1200 Switzen Avenue St. Louis, MO 63147	Varco-Floc
W. E. Zimmie, Inc. 810 Sharon Drive Westlake, OH 44145	Zimmite

10.7.3 Other Chemicals

A large number of manufacturers generate sodium hydroxide; it is produced concurrently with chlorine, and sometimes elemental sodium. The compound is produced at nearly 100 plant sites in this country, and bulk storage is available in most major cities. Designers are referred to chemical suppliers in their geographical area for this commodity.

Carbon dioxide is generated at some 75 locations around the country, and in addition the liquefied gas is stored in bulk in nearly all major cities. Some twenty manufacturers are involved in these operations, and designers are referred to local chemical suppliers for further information on carbon dioxide.

About ten manufacturers produce soda ash in the United States. The 15 plant sites involved are widely scattered geographically, as follows in Table 10-15:

TABLE 10-15

SODA ASH MANUFACTURERS

Ammonia-Soda Plants

Allied Chemical Corp., Syracuse, N.Y.; Detroit, Mich.; Baton Rouge, La.

Diamond Alkali Co., Painesville, Ohio

Olin Mathieson Chemical Corp., Saltville, Va.; Lake Charles, La.

Pittsburgh Plate Glass Co., Barberton, Ohio; Corpus Christi, Texas

Wyandotte Chemicals, Corp., Wyandotte, Mich.

Natural Soda Plants

American Potash and Chemical Corp., Trona, Calif.

FMC Corp., Green River, Wyoming

Pittsburgh Plate Glass Co., Bartlett, Calif.

Stauffer Chemical Co., Westend, Calif.; Green River, Wyoming

Carbonation of Caustic Liquors

Dow Chemical Co., Freeport, Texas

The preceding list is from a reference written to supply such information on these and many other related compounds:

Faith, W. L.; Keyes, D. B.;
& Clark, R. L., Industrial
Chemicals, 3rd Ed., John
Wiley & Son, New York (1965).

10.8 Chemical Feeders

Chemical feed systems must be flexibly designed to provide for a high degree of reliability in light of the many contingencies which may affect their operation. Thorough waste characterization in terms of flow extremes and chemical requirements should precede the design of the chemical feed system. The design of the chemical feed system must take into account the form of each chemical desired for feeding, the particular physical and chemical characteristics of the chemical, maximum waste flows and the reliability of the feeding devices.

In suspended and colloidal solids removal from wastewaters the chemicals employed are generally in liquid or solid form. Those in solid form are generally converted to solution or slurry form prior to introduction to the wastewater stream; however, some chemicals are fed in a dry form. In any case, some type of solids feeder is usually required. This type of feeder has numerous different forms due to wide ranges in chemical characteristics, feed rates and degree of accuracy required. Liquid feeding is somewhat more restrictive, depending mainly on liquid volume and viscosity.

The capacity of a chemical feed system is an important consideration in both storage and feeding. Storage capacity design must take into account the advantage of quantity purchase versus the disadvantage of construction cost and chemical deterioration with time. Potential delivery delays and chemical use rates are necessary factors in the total picture. Storage tanks or bins for solid chemicals must be designed with proper consideration of the angle of repose of the chemical and its necessary environmental requirements, such as temperature and humidity. Size and slope of feeding lines are important along with their materials of construction with respect to the corrosiveness of the chemicals.

Chemical feeders must accommodate the minimum and maximum feeding rates required. Baker (12) indicates that manually controlled feeders have a common range of 20:1, but this range can be increased to about 100:1 with dual control systems. Chemical feeder control can be manual, automatically proportioned to flow, dependent on some form of process feedback, or a combination of any two of these. More sophisticated control systems are feasible if proper sensors are available. If manual control systems are specified with the possibility of future automation, the feeders selected should be amenable to this conversion with a minimum of expense. An example would be a feeder with an external motor which could easily be replaced with a variable speed motor or drive when automation is installed (12). Standby or backup units should be included for each type of feeder used. Reliability calculations will be necessary in larger plants with a greater multiplicity of these units. Points of chemical addition and piping to them should be capable of handling all possible changes in dosing patterns in order to have proper flexibility of operation. Designed flexibility in hoppers, tanks, chemical feeders and solution lines is the key to maximum benefits at least cost (12).

Liquid feeders are generally in the form of metering pumps or orifices. Usually these metering pumps are of the positive-displacement variety, plunger or diaphragm type. The choice of liquid feeder is highly dependent on the viscosity, corrosivity, solubility, suction and discharge heads, and internal pressure-relief requirements (13). Some examples are shown in Figure 10-13 and Figure 10-14. In some cases control valves and rotameters may be all that is required. In other cases, such as lime slurry feeding, centrifugal pumps with open impellers are used with appropriate controls (14). More complete descriptions of liquid feeder requirements can be found in the literature and elsewhere (13).

Solids characteristics vary to a great degree and the choice of feeder must be considered carefully, particularly in the smaller-sized facility where a single feeder may be used for more than one chemical. Generally, provisions should be made to keep all chemicals cool and dry. Dryness is very important, as hygroscopic (water absorbing) chemicals may become lumpy, viscous or even rock hard; other chemicals with less affinity for water may become sticky from moisture on the particulate surfaces, causing increased arching in hoppers. In either case, moisture will affect the density of the chemical and may result in under-feed. Dust removal equipment should be used at shoveling locations, bucket elevators, hoppers and feeders for neatness, corrosion prevention and safety reasons. Collected chemical dust may often be used.

The simplest method for feeding solid chemicals is by hand. Chemicals may be preweighed or simply shoveled or poured by the bagful into a dissolving tank. This method is of economic necessity limited to very small operations, or to chemicals used in very weak solutions.

Because of the many factors, such as moisture content, different grades and compressibility, which can affect chemical density (weight to volume ratio), volumetric feeding of solids is normally restricted to smaller plants, specific types of chemicals which are reliably constant in composition and low rates of feed. Within these restrictions, several volumetric types are available. Accuracy of feed is usually limited to plus or minus 2% by weight but may be as high as plus or minus 5%.

FIGURE 10-13

PLUNGER TYPE METERING PUMP

(COURTESY OF WALLACE & TIERNAN)

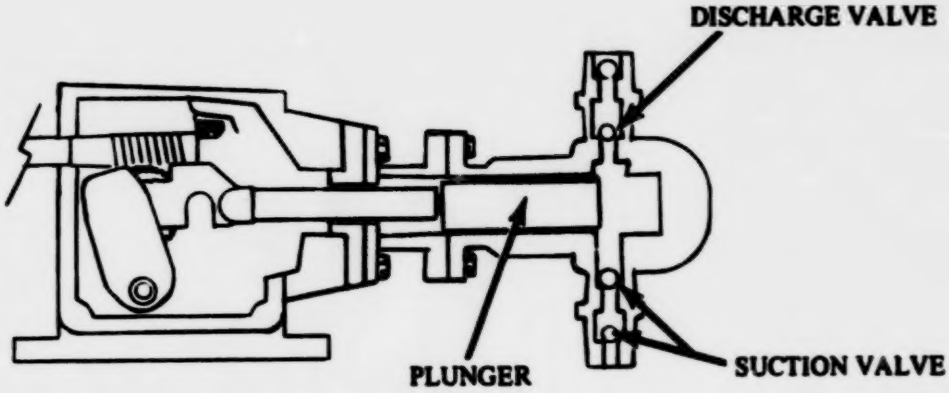
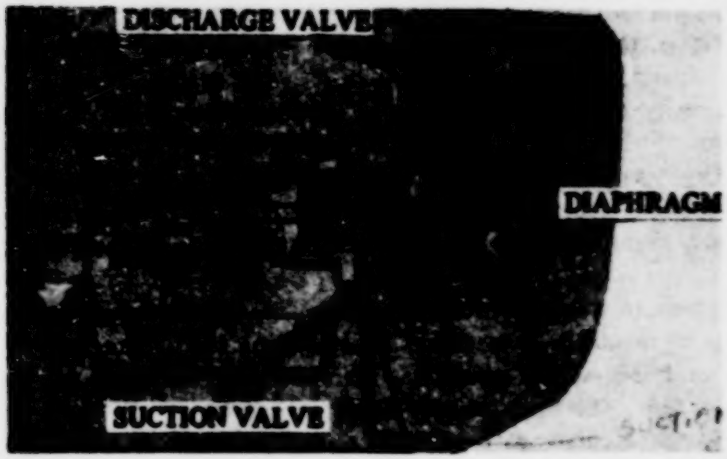

DISCHARGE VALVE
PLUNGER
SUCTION VALVE

FIGURE 10-14

DIAPHRAGM TYPE METERING PUMP

(COURTESY OF WALLACE & TIERNAN)


DISCHARGE VALVE
DIAPHRAGM
SUCTION VALVE

One type of volumetric dry feeder uses a continuous belt of specific width moving from under the hopper to the dissolving tank. A mechanical gate mechanism regulates the depth of material on the belt, and the rate of feed is governed by the speed of the belt and/or the height of the gate opening. The hopper normally is equipped with a vibratory mechanism to reduce arching. This type of feeder is not suited for easily fluidized materials.

Another type employs a screw or helix from the bottom of the hopper through a tube opening slightly larger than the diameter of the screw or helix. Rate of feed is governed by the speed of screw or helix rotation. Some screw-type designs are self-cleaning, while others are subject to clogging. Figure 10-15 shows a typical screw-feeder.

Most remaining types of volumetric feeders generally fall into the positive-displacement category. All designs incorporate some form of moving cavity of a specific or variable size. In operation, the chemical falls by gravity into the cavity and is more or less fully enclosed and separated from the hopper's feed. The size of the cavity, and the rate at which the cavity moves and is discharged, governs the amount of material fed. The positive control of the chemical may place a low limit on rates of feed. One unique design is the progressive cavity metering pump, a non-reciprocating type. Positive-displacement feeders often utilize air injection to enhance flowability of the material. Some examples of positive-displacement units are illustrated in Figure 10-16.

The basic drawback of volumetric feeder design, i.e., its inability to compensate for changes in the density of materials, is overcome by modifying the volumetric design to include a gravimetric or loss-in-weight controller. This modification allows for weighing of the material as it is fed. The beam balance types measure the actual mass of material. This is considerably more accurate, particularly over a long period of time, than the less common spring-loaded gravimetric designs. Gravimetric feeders are used where feed accuracy of about plus or minus 1% is required for economy, as in large-scale operations and for materials which are used in small, precise quantities. It should be noted, however, that even gravimetric feeders cannot compensate for weight added to the chemical by excess moisture. Many volumetric feeders may be converted to loss-in-weight function by placing the entire feeder on a platform scale which is tared to neutralize the weight of the feeder.

Good housekeeping and need for accurate feed rates dictate that the gravimetric feeder be shut down and thoroughly cleaned on a regular basis. Although many of these feeders have automatic or semi-automatic devices which compensate to some degree for accumulated solids on the weighing mechanism, accuracy is affected, particularly on humid days when hygroscopic materials are fed. In some cases, built-up chemicals can actually jam the equipment.

FIGURE 10-15

SCREW FEEDER

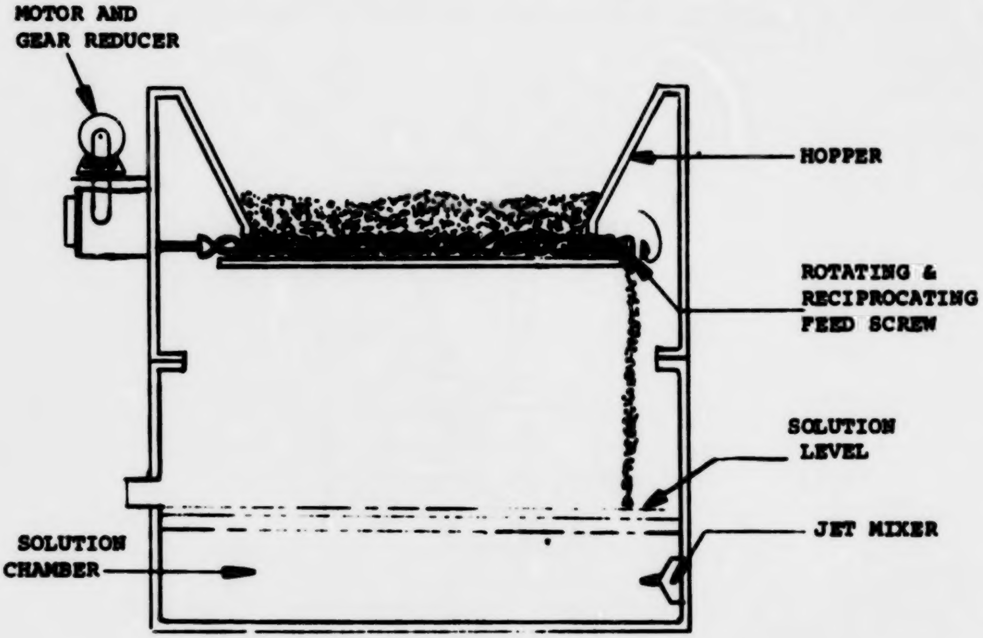
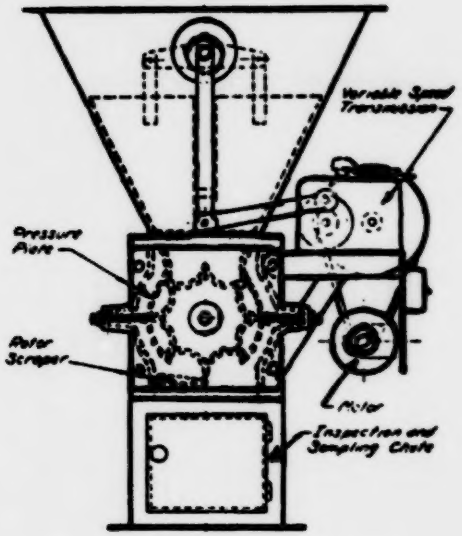

MOTOR AND
GEAR REDUCER
HOPPER
ROTATING &
RECIPROCATING
FEED SCREW
SOLUTION
LEVEL
JET MIXER
SOLUTION
CHAMBER

FIGURE 10-16

POSITIVE DISPLACEMENT SOLID FEEDER-ROTARY


Variable Speed Transmission
Pressure Plate
Rotor Scraper
Motor
Inspection and Sampling Chute

No discussion of feeders is complete without at least passing reference to dissolvers, as any metered material must be mixed with water to provide a chemical solution of desired strength. Most feeders, regardless of type, discharge their material to a small dissolving tank which is equipped with a nozzle system and/or mechanical agitator depending on the solubility of the chemical being fed. Solid materials, such as polyelectrolytes, may be carefully spread into a vortex spray or washdown jet of water immediately before entering the dissolver. It is essential that the surface of each particle become thoroughly wetted before entering the feed tank to ensure accurate dispersal and to avoid clumping, settling or floating.

A dissolver for a dry chemical feeder is unlike a chemical feeding mechanism, which by simple adjustment and change of speed can vary its output tenfold. The dissolver must be designed for the job to be done. A dissolver suitable for a rate of 10 lb/hr may not be suitable for dissolving at a rate of 100 lb/hr. As a general rule, dissolvers may b oversized, but dissolvers for commercial ferric sulfate or lime slakers do not perform well if greatly oversized.

It is essential that specifications for dry chemical feeders include specifications on dissolver capacity. A number of factors need to be considered in designing dissolvers of proper capacity. These include detention times and water requirements, as well as other factors specific to individual chemicals.

The capacity of a dissolver is based on detention time, which is directly related to the wettability of rate of solution of the chemical. Therefore, the dissolver must be large enough to provide the necessary detention for both the chemical and the water at the maximum rate of feed. At lower rates of feed, the strength of solution or suspension leaving the dissolver will be less, but the detention time will be approximately the same unless the water supply to the dissolver is reduced. When the water supply to any dissolver is controlled for the purpose of forming a constant strength solution, mixing within the dissolver must be accomplished by mechanical means, because sufficient power will not be available from the mixing jets at low rates of flow. Hot water dissolvers are also available in order to minimize the required tankage.

The foregoing descriptions give some indication of the wide variety of materials which may be handled. Because of this variety, a modern facility may contain any number of a variety of feeders with combined or multiple materials capability. Ancillary equipment to the feeder also varied according to the material to be handled. Liquid feeders encompass a limited number of design principles which account for density and viscosity ranges. Solids feeders, relatively speaking, vary considerably due to the wide range of physical and chemical characteristics, feed rates and the degree of precision and repeatability required.

TABLE 10-16

TYPES OF CHEMICAL FEEDERS

Type of feeder	Use	Limitations	
		Capacity (cubic feet per hr)	Range
Dry feeder:			
Volumetric:			
Oscillating plate	Any material, granules or powder.	0.01 to 35	40 to 1
Oscillating throat (universal)	Any material, any particle size.	0.002 to 100	40 to 1
Rotating disc	Most materials including NaF, granules or powder.	0.01 to 1.0	20 to 1
Rotating cylinder (star)	Any material, granules or powder.	8 to 2,000 or 7.2 to 300	10 to 1 or 100 to 1
Screw	Dry, free flowing material, powder or granular.	0.05 to 18	20 to 1
Ribbon	Dry, free flowing material, powder, granular, or lumps.	0.002 to 0.16	10 to 1
Belt	Dry, free flowing material up to 1½-inch size, powder or granular.	0.1 to 3,000	10 to 1 to 100 to 1
Gravimetric:			
Continuous–belt and scale	Dry, free flowing, granular material, or floodable material.	0.02 to 2	100 to 1
Loss in weight	Most materials, powder granular or lumps.	0.02 to 80	100 to 1
Solution feeder:			
Nonpositive displacement:			
Decanter (lowering pipe)	Most solutions or light slurries.	0.01 to 10	100 to 1
Orifice	Most solutions	0.16 to 5	10 to 1
Rotameter (calibrated valve)	Clear solutions	0.005 to 0.16 or 0.01 to 20	10 to 1
Loss in weight (tank with control valve).	Most solutions	0.002 to 0.20	30 to 1
Positive displacement:			
Rotating dipper	Most solutions or slurries	0.1 to 30	100 to 1
Preportioning pump:			
Diaphragm	Most solutions. Special unit for 5% slurries.[1]	0.004 to 0.15	100 to 1
Piston	Most solutions, light slurries.	0.01 to 170	20 to 1
Gas feeders:			
Solution feed	Chlorine	8,000 lb/day max	20 to 1
	Carbon dioxide	6,000 lb/day max	20 to 1
Direct feed	Chlorine	300 lb/day max	10 to 1

[1] Use special heads and valves for slurries.

Table 10-16 describes several types of chemical feeders commonly used in wastewater treatment.

10.9 Additional Reading

10.9.1 General

1. *Water Quality and Treatment,* Third Edition, American Water Works Association, Inc., McGraw-Hill, NY (1971).

2. *Water Treatment Plant Design,* American Society of Civil Engineers, American Water Works Association, and Conference of State Sanitary Engineers, NY (1969).

3. Ockershausen, R. W., "Safe Handling of Water Works Chemicals," *JAWWA,* 63:6, p. 336 (1971).

4. "Chemicals Used in Treatment of Water and Wastewater," BIF Reference No. 1.21-15, BIF Industries, Providence, RI (1970).

10.9.2 Aluminum Compounds

5. Truitt, V., "We Converted to Liquid Alum Treatment," *Public Works,* 99:11, p. 88 (1968).

6. "Aluminum Sulfate," Allied Chemical Corporation, Morristown, NJ (1968).

7. "Alum," American Cyanamid Company, Wayne, NJ (1964).

8. "Aluminum Sulfate," Stauffer Chemical Company, NY (1967).

10.9.3 Iron Compounds

9. "Handling Ferric Chloride," Dow Chemical Company, Midland, MI (1965).

10. "Pennsalt Ferric Chloride," Pennsalt Chemicals Corporation, Philadelphia, PA (1965).

10.9.4 Lime

11. "Chemical Lime Facts," National Lime Association, Washington, D.C. (1964).

12. Carr, R. L., "Limestone to Lime to Slaked Lime," *Water and Sewage Works*, Reference No. 113:R.N., p. R-61 (1966).

13. "Chemicals Used in Water and Wastewater Treatment - Calcium Hydroxide (Hydrated Lime)," *Water and Wastes Engineering*, 7:1, p. 53 (1970).

14. "Chemicals Used in Water and Wastewater Treatment - Quicklime," *Water and Wastes Engineering*, 7:5, p. 52 (1970).

15. "Lime Handling, Application & Storage in Treatment Processes," National Lime Association, Washington, D. C. (1971).

16. "Chemistry and Technology of Lime and Limestone," Boynton, R. S., John Wiley & Sons, New York, NY (1966).

10.9.5 Soda Ash

17. "Soda Ash," Allied Chemical Corporation, NY (1966).

18. "What's Different About FMC Soda Ash?", FMC Corporation, NY (1964).

19. "Stauffer 'Natural' Soda Ash," Stauffer Chemical Company, NY (1964).

20. "Wyandotte Soda Ash," Wyandotte Chemicals Corporation, Wyandotte, Michigan (1955).

10.9.6 Caustic Soda

21. "Caustic Soda," PPG Industries, Inc., Chemical Division, Pittsburgh, PA (1969).

22. "Unloading Liquid Caustic from Tank Cars," Manufacturing Chemists Association, Washington, D.C. (1968).

23. "Properties and Essential Information for Safe Handling and Use of Caustic Soda," Manufacturing Chemists Association, Washington, D.C. (1968).

24. "Caustic Soda," Allied Chemical Corporation, NY (1963).

25. "Caustic Soda Handbook," Diamond Alkali Company, Cleveland, OH (1967).

26. "Dow Caustic Soda Handbook," Dow Chemical Company, Midland, MI (Undated).

27. "Hooker Caustic Soda," Hooker Chemical Corporation, Niagara Falls, NY (1966).

28. "Olin Caustic Soda," Olin Mathieson Chemical Corporation, NY (1961).

29. "Pennsalt Caustic Soda," Pennsalt Chemicals Corporation, Philadelphia, PA (1964).

30. "Caustic Soda," Stauffer Chemical Company, NY (1965).

31. "Caustic Soda by Wyandotte," Wyandotte Chemicals Corporation, Wyandotte, MI (1961).

10.10 References

1. "Chemicals Used in Water and Wastewater Treatment-Aluminum Sulfate," *Water and Wastes Engineering*, Dec. 6:12, p. 46 (1969).

2. "Chemicals Used in Water and Wastewater Treatment-Ferric Chloride," *Water and Wastes Engineering*, 7:3, p. 65 (1970).

3. "Ferri-Floc for Water and Wastewater Treatment," Cities Service Co., Atlanta, GA (1972).

4. Schworm, W. B., "Iron Salts for Water and Waste Treatment," *Public Works*, 94:10, p. 118 (1963).

5. "Lime for Water and Wastewater Treatment," BIF Reference No. 1.21-24, BIF Industries, Providence, RI (June, 1969).

6. National Lime Association, direct communication (May, 1971).

7. "Caustic Soda," PPG Industries, Inc., Chemical Division, Pittsburgh, PA (1969).

8. Haney, P. D. and Hamann, C. L., "Recarbonation and Liquid Carbon Dioxide," *JAWWA*, 61:10, p. 512 (1969).

9. Culp, R. L. and Culp, G. L., *Advanced Wastewater Treatment*, Van Nostrand-Reinhold Company, NY (1971).

10. Carr, R. L., "Polyelectrolyte Coagulant Aids - Dry and Liquid Handling and Application," *Water and Sewage Works*, Reference No. 114:R.N., p. 4-64 (1967).

11. Russo, F. and Carr, R. L., "Polyelectrolyte Coagulant Aids and Flocculants: Dry and Liquid, Handling and Application," *Water and Sewage Works*, Reference No. 117:R.N., p. R-72 (1970).

12. Baker, R. J., "Chemical Feed Systems Determine Plant Efficiency and Reliability," *Water and Sewage Works*, Reference No. 116:R-21 (November, 1969).

13. Russo, F., and Carr, R. L., Jr., "Polyelectrolyte Coagulant Aids and Flocculants: Dry and Liquid, Handling and Application," *Water and Sewage Works*, Reference No. 117:R-72 (November, 1970).

14. Culp, R. L., and Culp, G. L., *Advanced Wastewatewater Treatment*, Van Nostrand Reinhold Company, NY (1971).

15. R. P. Lowe, "Chemical Feed Systems," 10th Annual Water Conf. of Eng. Soc. of W. PA (October 17-19, 1949).

CHAPTER 11

SLUDGE HANDLING AND DISPOSAL

11.1 Introduction

A high proportion of wastewater treatment capital and operating cost is due to sludge handling and disposal. Consequently, it is important that design engineers have reliable information on quantities and dewatering properties of sludges so that solids handling facilities can be properly designed.

Information on sludges produced in conventional wastewater treatment is widely scattered, and results from different locations are not related well to each other. There is even less information available on sludges produced by the newer processes for phosphorus removal.

There has recently been a concerted effort to provide this necessary information. One group involved is the staff of the Ultimate Disposal Research Program operating as part of the Advanced Waste Treatment Research Laboratory of the National Environmental Research Center in Cincinnati.

Material in this chapter comes from information compiled and organized by that group. Their basic data were obtained in several ways:

a. Grants and contracts aimed at obtaining comparative information on phosphorus-removal sludges have been funded.

b. Reports have been prepared which summarize information obtained on sludges in EPA Research and Development Grants where useful sludge data have been collected.

c. Field crews have visited certain plants where promising processes are being implemented and have obtained comparative data on sludges involved.

d. Visits have been made to municipalities and sludge data have been obtained from plant staffs.

e. Published literature has been surveyed and information of value on sludges has been extracted.

f. Memoranda have been prepared to organize the information available and to respond to specific needs.

g. Experimental investigations of sludge dewatering are being conducted at several locations by Advanced Waste Treatment Laboratory staff.

All of these routes are still being actively pursued, although the collection of information is far from complete. However, there was such urgent need for this information (e.g., inclusion in this manual) that an interim report was prepared, and this chapter was developed from it.

Progress on sludge handling and disposal technology will be dealt with in detail in a forthcoming design manual on the subject (1).

11.2 Generalized Assessment of Filtration Rates

At least two groups (2) (3) have developed cost information computer programs which can be used to predict the cost of alternative processing schemes for wastewater treatment and for phosphorus removal processes. These programs have included in their inherent logic (not as input data) typical values for dewatering rates, normal sludge mass, and additional sludge produced by phosphorus removal processes.

Aside from the output from these cost estimated computer programs, reliable estimates of the dewaterability of various sludges can be of substantial value to those concerned with process design. That type of information is presented in this chapter.

When attempting to decide on proper values for sludge dewatering properties, there is a statistical problem of within-plant variations and between-plant variations. The between-plant variations are sometimes extreme. This is illustrated in the following hypothetical compilation:

TABLE 11-1

HYPOTHETICAL COMPILATION OF DEWATERING RATES

	Vacuum Filter Yield (lb/hr-sq ft)	
	Avg. (40 Plants)	Std. Deviation
Digested Mixed Sludge	4.0	2.0
Raw Mixed Sludge	5.0	3.0

The designer wishing to use data of this type to predict within-plant differences between raw and digested sludges will find tabulated values are of little value because standard deviations are so great. Within-plant variations between raw and digested sludge are less than indicated in the table. Unfortunately, within-plant comparisons are virtually unavailable because usually only one type of sludge at a given plant is dewatered.

Information relating various within-plant sludges to each other should eventually be collected. In this chapter there has been an attempt to estimate such relationships based on experience and information from the sources previously cited. Relative values are suggested here. These generalized figures should be used only when there is no other information available.

The information in this section relates only to vacuum filtration. The important area of centrifugation has been introduced in other chapters, when such information was available. Section 11.4 also describes some case studies in which centrifugation was used as a dewatering means, but data are insufficient for development of generalized design parameters.

There has been no consideration given to lime sludges in this introductory section. Sludges produced by lime addition, even when added to the primary clarifier, contain large quantities of calcium carbonate and filter better than conventional or alum and iron sludges. In addition, a different approach is needed than for alum, iron, and conventional sludges. Section 11.4 describes case studies where lime sludge is dewatered by various means, and Section 11.3.2 presents fundamentals involved.

11.2.1 Conventional Sludges

The digested total sludge from a conventional primary plus activated sludge plant is probably the commonest sludge encountered in wastewater treatment. Consequently, it has been selected as the standard against which other sludges are compared.

Table 11-2 shows estimated relative filtration yields and conditioning costs of various wastewater sludges compared to a standard digested primary plus waste activated sludge. These values are for belt-type vacuum filters operated at optimum conditions, preceded by a reasonable processing sequence including thickening if clarifier sludge is dilute. Filter cloth selection was appropriate for the situation to get satisfactory solids capture.

TABLE 11-2

DESIGN FACTORS FOR FILTRATION OF CONVENTIONAL SLUDGES

Sludge Type	Relative* Yield	Relative Cost of Conditioning**	Cake Solids Content
Raw Primary Plus WAS	1.4	0.7	28
Digested Primary Plus WAS	1.0	1.0	26
Raw Primary Plus TF	1.8	0.6	32
Digested Primary Plus TF	1.3	0.9	30
Digested Primary	2.3	0.7	33
Raw Primary	3.2	0.5	35

*Avg. yield for digested primary plus WAS is estimated at 4.0 lb/hr-sq ft

**Average cost of conditioning is $10/dry ton, 1974 costs

An average value of 4.0 lb/hr-sq ft is suggested for the yield when filtering digested primary plus waste activated sludge. If a good estimate of the yield for this or any other sludge is available, it should be utilized and estimates of the yields for the other sludges obtained by means of the factors in Table 11-2.

Mass of sludge produced in wastewater treatment can be estimated by means of well-established procedures discussed in Section 11.3. Reduction in mass by digestion can be estimated by volatile solids reduction. Since digestion reduces mass approximately 50 percent, required filter area will be less for digested sludge than for raw sludge despite slightly higher yields for raw sludges.

Some shortcomings of the tabulated numbers are evident. No influence of solids content is indicated, although it is well known that, up to about 8% solids, filter yield is proportional to solids content. It must be presumed that appropriately designed thickeners are used. It becomes clear that when a filter yield is chosen, there are implicit assumptions about the operations upstream.

Effect of elutriation is not included in Table 11-2. Elutriation would reduce lime demand substantially. Polymer required for conditioning would be reduced as well; elutriation is virtually a necessity for some digested sludges (4).

No indication of range in solids content of filtered sludge cake is given. This is of importance when the sludge is to be incinerated. Some data (5) indicate that digested primary and waste activated sludge would have a solids content of about 28%. For raw primary and waste activated sludge, it would be about 30%. Slightly lower values are suggested in Table 11-2.

11.2.2 Iron and Alum Sludges

Sludges produced when adding ferric or aluminum salts have a substantially different character than conventional sludges. Quantities are greater and handling and dewatering properties are different. Changing the addition point in the processing scheme also produces changes in the character of the sludge.

Table 11-3 compares estimated vacuum filtration yields and conditioning costs of the various metal-effected sludges relative to conventional digested primary plus waste activated sludge.

TABLE 11-3

DESIGN FACTORS FOR FILTRATION
OF CONVENTIONAL PLUS METAL SLUDGES

Sludge Type	Relative Yield*	Relative Cost of Conditioning*
Aluminum to Primary		
Raw Primary Plus WAS	1.2	1.3
Digested Primary Plus WAS	0.85	1.4
Raw Primary Plus TF	1.35	1.2
Digested Primary Plus TF	1.0	1.3
Digested Primary	1.05	1.0
Raw Primary	1.5	0.9
Aluminum to Aerator		
Raw Primary Plus WAS	1.2	1.3
Digested Primary Plus WAS	0.85	1.4

Iron to Primary		
Raw Primary Plus WAS	1.3	1.1
Digested Primary Plus WAS	0.95	1.2
Raw Primary Plus TF	1.5	1.0
Digested Primary Plus TF	1.1	1.1
Digested Primary	1.2	0.8
Raw Primary	1.6	0.7
Iron to Aerator		
Raw Primary Plus WAS	1.3	1.1
Digested Primary Plus WAS	0.95	1.2

*Yields and costs are related to yield and cost obtained with digested primary plus WAS when no chemicals are added to the wastewater. See Table 11-2.

Estimates will be more accurate if absolute yields for one or more sludges can be determined for a particular situation. The factors in the table can then be used to predict other yields. If no information is available, filter yield for digested conventional primary plus waste activated sludge may be estimated to be 4.0 lb/hr-sq ft.

11.3 Quantity of Phosphorus Removal Sludges

11.3.1 Iron and Alum Sludges

When ferric iron or trivalent aluminum are added to wastewater containing phosphorus, a metal phosphate precipitates and excess metal forms an insoluble hydroxide. The hydroxide acts as a flocculant precipitate which carries with it other suspended solids when it settles. In addition, soluble organic material apparently is adsorbed on the precipitate and removed, as discussed later.

The amount and nature of additional sludge formed is a function of the process addition point. For example, addition to the primary clarifier causes removal of much more suspended solids than occurs in conventional sedimentation. Also, soluble organic material is removed there. However, the biological stage which follows receives a lower BOD wastewater; consequently, the amount of biological sludge is reduced correspondingly.

There are available tabulations which present average observed values of the amount of sludge typically produced when iron and aluminum are added to wastewater (6). However, there is still need for a more precise method of predicting sludge production so designers can more accurately size sludge processing equipment.

A detailed procedure is presented here for the case where these metals are added to the conventional processing sequence of primary treatment plus either activated sludge or trickling filter. In this procedure the chemicals may be added in primary or secondary treatment, or as a tertiary process. The method requires that the following information be available: complete processing sequence, wastewater characteristics (e.g., SS, BOD, soluble TOC), and efficiency of solids removal in clarifiers. Methods are outlined for predicting the following: chemical sludge produced, extra sludge from improved clarifier efficiency, extra sludge from removal of soluble materials by the chemicals, and reduced amount of biological sludge when biological plant load is reduced. In this manner, total sludge production, related to processing conditions and wastewater quality, can be determined. The method is given in Sections 11.3.1.1 through 11.3.1.5.

11.3.1.1 Theoretical Production of Chemical Sludge

Chemical sludge production may be calculated by assuming the metal added reacts first with phosphorus compounds which are thereby removed, and that any excess metal forms hydroxide. The calculation is illustrated in Table 11-4.

TABLE 11-4

CALCULATION OF CHEMICAL SLUDGE PRODUCTION

Given: P_{in} = 10 mg/l

P_{out} = 1 mg/l

Al Case: dose of 19 mg/l Al
(Al/P atomic ratio = 2.18)

Fe Case: dose of 40 mg/l Fe
(Fe/P atomic ratio = 2.22)

Aluminum

Stoichiometry:

$Al + PO_4 = Al\ PO_4$

$Al + 3\ OH = Al\ (OH)_3$

(9 mg/l P)/(31) = 0.29 millimoles/l Al PO_4
(M.W. = 122)

(19 mg/l Al)/27 = 0.70 mmol/l Al

0.70 - 0.29 = 0.41 mmol/l Al excess, goes to Al $(OH)_3$
(MW = 78)

(0.29 mmol/l Al PO_4)(122) = 35.4 mg/l Al PO_4

(0.41 mmol/l Al $(OH)_3$)(78) = 32.0 mg/l Al $(OH)_3$

Total = 35.4 + 32.0 = 67.4 mg/l

Sludge Produced:

67.4 mg chemical sludge per liter of wastewater treated
or 67.4/19 = 3.55 mg chemical sludge per mg Al
or 67.4/9 = 7.48 mg chemical sludge per mg P removed

Iron

Stoichiometry:

$Fe + PO_4 = Fe\ PO_4$

$Fe + 3\ OH = Fe\ (OH)_3$

(9 mg/l P)/(31) = 0.29 mmol/l Fe PO_4
(M.W. = 150.8)

(40 mg/l Fe)/(55.84) = 0.72 mmol/l Fe

0.72 - 0.29 = 0.43 mmol/l Fe excess, goes to $Fe(OH)_3$
(M.W. = 106.8)

(0.29 mmol/l Fe PO_4)(150.8) = 43.7 mg/l Fe PO_4

(0.43 mmol/l Fe $(OH)_3$)(106.8) = 45.9 mg/l Fe $(OH)_2$

Total = 43.7 + 45.9 = 89.6 mg/l

Sludge Produced:

89.6 mg of chemical sludge per liter of wastewater treated
or 89.6/40 = 2.13 mg chemical sludge per mg Fe
or 89.6/9 = 9.96 mg chemical sludge per mg P removed

The illustration in Table 11-4 shows that at approximately the same atomic ratio of metal to phosphorus, 48 percent more chemical sludge is predicted for iron addition than for aluminum addition.

Sludge production is related in Table 11-4 to mass of phosphorus removed per liter of sewage treated. This is the most rational procedure when sludge mass is calculated. However, it is frequently convenient to relate sludge mass to the mass of metal added. The following are estimated values for the ratio of sludge mass to metal dose as the metal/P atomic ratio increases from zero to infinity:

Al:	4.52 to 2.89 mg sludge/mg Al
Fe:	2.70 to 1.92 mg sludge/mg Fe

Information to verify these calculated values is scarce. Swedish experience (7) indicates the additional weight of sludge may be calculated from the following relationships:

sludge (mg/l of wastewater) = (4) (Al dose, mg/l)

sludge (mg/l of wastewater) = (2.5) (Fe dose, mg/l)

That reference does not elaborate on the source of information although it was drawn from Swedish practice at the time. It is not stated whether additional mass from improved suspended and dissolved solids removal is included. The great majority of plants in Sweden were adding chemicals as a post-treatment after biological treatment. Consequently, it may be presumed that these figures indicate total increase in sludge mass from Al or Fe added to the final effluent from a biological treatment plant.

Other studies (8) estimated additional sludge produced by alum addition to the activated sludge process. They report 4 to 3 mg addition solids per mg of Al as dose level increased from 10 to 20 mg Al/1.

Another report (9) has estimated chemical sludge production when alum was added to secondary effluents. The following relationship was observed:

chemical sludge (mg/l of wastewater) = (5.2)(Al dose, mg/l)

Al/P atomic ratio range: 2.0 to 2.8

This is about 20 percent higher than the Swedish figure and about 45 percent higher than indicated in Table 11-4 for the stoichiometric relationship.

Data available, then, indicate chemical sludge mass higher than the values represented by the simple stoichiometry indicated in Table 11-4. There is reason to conclude that chemical sludge production could be higher than stoichiometric values because other cations (such as heavy metals) and anions can be incorporated into the complex ligand structures of aluminum and ferric hydroxides. Consequently, it is recommended that with either Al or Fe, sludge mass be estimated to be 35 percent higher than the simple stoichiometric values.

11.3.1.2 Theoretical Production of Biological Sludge

Biological sludge is produced and wasted in both the activated sludge process and the trickling filter process.

One source (10) reports the following equation for excess sludge production in the conventional activated sludge process:

$$W_v/M = (0.79\ F/M) - (0.0714)$$

where W_v = lb volatile suspended solids produced per day

M = lb mixed liquor volatile suspended solids in the aerator

F = lb 5-day BOD removed per day

Rearranging this equation,

$$W_v/F = (0.79) - (0.0714/F/M)$$

The equation now gives the mass of volatile solids produced per unit mass of BOD removed (W_v/F) as a function of the food to mass ratio.

A recent report (11) states that 50 percent of the sum of the BOD and suspended solids fed to a trickling filter is converted to waste sludge.

If suspended solids and BOD concentrations in the wastewater as it enters the biological treatment step are known, sludge production can be estimated by means of the above relationships.

11.3.1.3 Increase in Production of Sludge From Sewage Solids

Suspended solids and dissolved solids in wastewater are removed by chemical treatment. Sludge mass is increased by the direct contribution of the mass of these removed solids.

One report (12) has summarized data on suspended solids removal from well designed physical-chemical pilot plants. For an average sewage and effective clarifiers, suspended solids removal of 85 percent can be achieved. If the sewage is high in suspended solids, removals will be better. This figure is substantially higher than the 50-75% removal range suggested previously (13). However, those earlier figures included plants where chemicals were added to an existing facility which might have been overloaded or were functioning poorly for other reasons.

The mass of sludge produced from suspended solids removed is calculated directly by multiplying the change in suspended solids concentration by the volume of sewage.

There is also evidence (12) which indicates an appreciable mass of dissolved materials (i.e., filtrable through a 0.45 micron membrane filter) is removed by chemical treatment. The following table presents this information along with data collected from other sources.

TABLE 11-5

REMOVAL OF DISSOLVED MATERIALS DURING CHEMICAL ADDITION

Source	Percent Removals			
	Fe		Al	
	SCOD	STOC	SCOD	STOC
Eimco				
pilot plant (14)			40	30
laboratory (14)				ca. 30
Friedman et al (15)		30		
Farrell (16)				27
Westrick and Cohen (17)	45	42		

Note: SCOD is soluble COD
STOC is soluble TOC

Some of that date (16) represent the average removal obtained in seven batch experiments in which raw sewage was treated with alum in a small clarifier (150 liter batches). Al/P atomic ratio ranged from 1.6 to 2.8 and removals from 20 to 32 percent.

Based on the above data, it is concluded that soluble TOC removals will average 30 percent. Data are scant for COD and none appear available for BOD. It is assumed, however, that removals will be the same for STOC.

The mass of sludge associated with the soluble TOC removed is more than the carbon content involved. It is estimated that, on the average, the soluble compounds removed by Al and Fe treatment will average 40 percent carbon on a volatile solids basis. This figure is in accord with the carbon content of compounds tested in developing the TOC test for wastewaters (18). Thus, to obtain an estimate of the mass of sludge (volatile basis) produced, TOC removed should be multiplied by 1/0.40 or 2.5.

If only soluble COD removed or soluble BOD removed is known, the problem is more difficult, primarily because the relationship between mass and COD or BOD changes as the degree of oxidation of the sewage changes. For a secondary effluent, volatile mass/COD is generally considered to be about 1.5 (19). Assuming the BOD is about 70 percent of the COD, volatile mass/BOD is 2.15. For raw sewage, which has a low degree of oxidation, the COD is expected to be higher for a given mass. Volatile mass/COD for a raw sewage is estimated to be 1.1, approximately 10 percent lower than lowest values previously found (19) for secondary effluent. Volatile mass/BOD would be about 1.6.

In the above discussion, mass relates to mass of volatile solids. To obtain total mass, non-volatile solids must be included. It is estimated that the ash content of these soluble solids is about 15 percent. Mass of volatile solids should be multiplied by 100/(100-15) or 1.18 to get total solids.

Calculation of sludge mass resulting from removal of soluble material will require information on the absolute value of soluble TOC, COD or BOD. These values are rarely measured. An examination of previous data (14), where both SCOD and STOC were measured, showed that ratios of soluble to total COD and TOC in the influent sewage were 0.30 and 0.38 respectively. Other data (16) showed an STOC/TOC ratio of 0.64. If data on soluble values of TOC or COD are not available, soluble TOC or COD may be estimated to be 50 percent of the total TOC or COD.

The correlations for predicting sludge production require information on total BOD removal. Some data (12) indicate total BOD, COD and TOC reductions are about the same and range from 50 to 85 percent depending on clarifier efficiency. Removal of

65 percent is thereby chosen as a representative figure and will be used in example calculations.

11.3.1.4 Summary of Calculation Procedures

The calculation procedure is described for various chemical addition points and is illustrated with a specific example in Section 11.3.1.5. Factors involved in that example are summarized here.

Chemicals Added to Primary

Chemical sludge is calculated on the basis of the phosphorus level and the chemical dose in the manner illustrated in Table 4. Estimated mass should be increased by about 35 percent (see Section 11.3.1.1).

Additional suspended solids removals in primary treatment are estimated using quantities derived per Section 11.3.1.3. The additional suspended solids mass removed should be added directly to total sludge mass.

Mass resulting from dissolved solids removal has to be estimated indirectly from soluble TOC, COD or BOD removal. Factors to be used in predicting mass have been discussed in Section 11.3.1.3. Equations relating soluble TOC, COD and BOD input levels to mass of soluble solids removed are summarized below:

$$\begin{aligned}(\text{soluble solids mg/l}) &= \text{STOC}_{in}(\text{mg/l})\text{x}0.30\text{x}2.5\text{x}1.18\\ &= \text{SCOD}_{in}(\text{mg/l})\text{x}0.30\text{x}1.1\text{x}1.18\\ &= \text{SBOD}_{in}(\text{mg/l})\text{x}0.30\text{x}1.6\text{x}1.18\end{aligned}$$

The last two equations apply only to influents before oxidative processes such as the activated sludge process.

Estimates of total TOC, COD or BOD removals are needed to estimate the loading on the biological processes which follow chemical treatment. However, this information is not needed for estimating sludge formation in the chemical treatment step.

Chemicals Added to Activated Sludge

The chemical sludge is calculated in the manner described in Section 11.3.1.1.

Addition of chemicals to the activated sludge process may produce a significant increase in BOD and suspended solids removals. This is particularly true of a high rate activated sludge plant or an overloaded plant. However, for a well-designed standard rate plant addition of chemicals will produce little additional sludge other than the chemical sludge itself. It is recommended that for a standard rate plant no increase in mass be included for additional suspended solids. For special cases such as an existing high rate plant, judgment will have to be made based on available information.

Chemicals Added After Biological Treatment

The chemical sludge is calculated as in Section 11.3.1.1. If the tertiary treatment follows a standard rate activated sludge plant, additional sludge from improved suspended solids and dissolved BOD removal can be neglected. For a trickling filter plant, however, additional mass removed could be important. For one plant (20) a substantial improvement in BOD and solids removal in a high rate trickling filter plant resulted from alum addition before the final clarifier. In one series of runs, overall BOD removals increased from 73% to 91% and suspended solids removal increased from 70% to 84%.

At another plant (21), under conditions where the final clarifier was not hydraulically overloaded, alum added after a standard rate trickling filter brought BOD and suspended solids from the 15-20 mg/l levels to below 10 mg/l.

These results can be used to calculate additional sludge brought down with the chemical sludge.

11.3.1.5 Illustrative Calculation

Sludge mass is calculated here for a plant utilizing primary treatment with aluminum introduced before clarification, followed by standard rate activated sludge. Details of the calculation are presented in Table 11-6. The existing plant is a properly designed plant operating within its design limits, which is changed only by introduction of a means to inject the chemical and flocculate it before primary clarification.

TABLE 11-6

CALCULATION OF TOTAL SLUDGE MASS WITH METAL ADDITION

Given: Plant utilizes primary clarification followed by standard rate activated sludge. Alum is to be added in a flocculation tank prior to the primary clarifier. Estimate the additional sludge produced by alum addition.

Primary clarifier of the existing plant is designed for 50% suspended solids removal (without chemical). Activated sludge plant is designed for an F/MLVSS ratio of 0.4.

Overall phosphorus removal is to be 90%. Estimate that this can be achieved at an Al/P atomic ratio of 2.0.

Influent to Primary: P = 10 mg/l
Total BOD = 172 mg/l
Suspended Solids = 240 mg/l
Soluble TOC = 40 mg/l

Sludge from Primary Treatment - No Chemicals

Estimate 50% SS removal, 30% total BOD removal and 0% STOC removal.

240 x 0.50 = 120 mg/l

Sludge from Primary Treatment - Alum Added

Estimate 90% P removal, 85% SS removal, 65% total BOD removal and 30% STOC removal.

Al Dose: (10/31)(2)(27) = 17.4 mg/l
10 x 0.90 = 9.0 mg/l P removal
9/31 = 0.29 meg (milliequivalent)/l Al PO_4
17.4/27 = 0.64 meg/l Al
0.64 - 0.29 = 0.35 meg/l Al excess
0.29 x 122 = 35.4 mg/l Al PO_4
0.35 x 78 = 27.3 mg/l Al $(OH)_3$
35.4 + 27.3 = 62.7 mg/l sludge produced

Chemical Solids: Increase by 35%, per Section 11.3.1.1
62.7 x 1.35 = 84.5 mg/l

Suspended Solids:	240 x 0.85	=	204 mg/l
Dissolved Solids:	40x0.30x2.5x1.18	=	35.4 mg/l
	Total Sludge:		323.9 mg/l

Sludge from Activated Sludge - No Chemicals

Equation from Section 11.3.1.2

$$W_v/F = (0.79) - (0.0714/F/M)$$

F/M is given as 0.4

$W_v/F = 0.79 - 0.18 = 0.61$ lb VSS/day/lb BOD removed per day

or 0.61 mg VSS/mg BOD removed

BOD in primary effluent is:

172 x 0.70 = 120 mg/l

BOD in final effluent is estimated to be almost 12 mg/l. See reference (6).

BOD removed = 120 - 12 = 108 mg/l

Therefore, 108 x 0.61 = 66 mg/l VSS produced

Assuming the sludge is 70% volatile,

66/0.70 = 94 mg/l waste sludge

Sludge from Activated Sludge - Al to Primary

BOD in primary effluent is:

172 x 0.35 = 60.2 mg/l

This is a very dilute waste. The existing activated sludge plant will be greatly oversized. Air requirements and the contact time may be reduced.

Keeping F/M the same (0.4).

W_v/F = 0.61 (see above)

Estimate BOD in final effluent to be 12 mg/l (see above).

BOD removed = 60.2 - 12 = 38.2 mg/l

Therefore, 38.2 x 0.61 = 23.3 mg/l VSS produced

Assuming sludge is 70% volatile,

23.3/0.70 = 33.3 mg/l waste sludge

Summary of Sludge Quantities Calculated

	Sludge Quantities (mg/l)	
	No Chemical	Al Added
Primary Clarifier		
Sludge from SS	120	204
Sludge from Dissolved Solids	0	35.4
Chemical Sludge	0	84.5
Total Clarifier Sludge	120	323.9
Secondary Clarifier		
Waste Activated Sludge	94	33.3
Total Sludge	214	357.2
Total Sludge Excepting Chemical Sludge	214	272.7

It is clear that chemical addition produces a substantial increase in the total sludge mass. The increase is greater than just the amount of chemical sludge because, in the no-chemical case, there is substantially more conversion of organic substances to carbon dioxide and water.

For a plant in which the chemical is added to the activated sludge process, the increase in mass will be virtually all due to the chemical sludge, since essentially the same amount of biological degradation is occurring as in the no-chemical case.

In the illustration, total sludge production (mass) was increased 67 percent when adding aluminum. The chemical sludge increased sludge mass 40 percent. The additional 27 percent sludge came from organic material which would have been oxidized had it reached the activated sludge aerator. If the chemical had been added to the aerator, sludge mass would have only increased 40 percent. The greater sludge mass which results when the chemical is added to the primary must be balanced against the reduced dimensions of the activated sludge aerator, the reduction in quantity of air needed, any differences in the dewatering properties of the different sludges, and differences in chemical demand which may be dependent on point of addition.

11.3.2 Lime Sludges

Lime is utilized for phosphorus removal in wastewater treatment by adding it just prior to primary clarification or after secondary clarification as a tertiary treatment process. As in the case of metal addition to wastewater, average values for the amount of sludge produced have been tabulated (6). However, there is need for more precise information applied to specific situations.

In response to this need, a procedure has been developed for predicting the mass of sludge produced when wastewater is treated with lime in a tertiary process. Details of this procedure are presented in this section. The method may be applied with minor modification when lime is added just prior to the primary clarifier.

Unlike metal addition, the amount of sludge produced by lime addition depends not only on the degree of phosphate removal desired, but also on alkalinity and other characteristics of the wastewater. In fact, both the amount of sludge produced and the degree of phosphorus removal achieved by raising the pH to a given level both depend on the nature of the water.

The procedure in predicting sludge quantity requires first that pH level be estimated for the desired phosphorus removal, then lime dose is estimated, and finally sludge mass can be calculated. The individual steps are listed below

a. From desired P-residual and knowledge of the water type, estimate pH level.

b. From pH level and alkalinity, estimate lime dose.

c. From amount of P removed, calculate the hydroxyapatite sludge produced.

d. From magnesium removal, calculate magnesium hydroxide sludge produced.

e. From calcium balance, calculate calcium carbonate sludge produced.

f. From pH and alkalinity, verify that sufficient carbonate is present to supply carbonate in sludge. If not, correct outlet calcium estimate.

g. Include in total sludge production the inerts introduced with the lime.

If lime is added prior to the primary settler, the procedure for estimating sludge for tertiary lime addition may be used with some modification. However, the correlation between pH and the ratio of lime dose to alkalinity was developed from data obtained by adding lime to secondary effluents. There is no assurance it would fully apply to raw wastewater. It would thus be desirable to establish lime dose needed by actual tests.

When lime is added prior to the primary settler, efficiency of the clarifier is increased substantially. From a previous review (12), it is estimated that for well designed plants, suspended solids removal efficiencies are in the order of 85 percent. This figure may be used to estimate additional sludge solids removed in the primary clarifier by lime addition.

The removal of dissolved organic solids by addition of lime prior to the primary appears to be less than obtained with metals. One study (14) reported very small or negative removals. Others (17) reported substantially lower removals with lime than with metal. Considering the relatively large mass of sludge from lime treatment, the small amount of dissolved organic solids brought down in primary treatment can be neglected.

Some prior data (14) have been utilized to estimate whether material balance calculations similar to those described above correlate with measured sludge production. This case is discussed in Section 11.4.2.2. Measured values were about 20 percent higher than the material balance figures. It is probably advisable to utilize a factor of safety of this magnitude in estimating chemical sludge production.

11.3.2.1 Estimation of Lime Dose and Sludge Production in Tertiary Treatment

A quantitative calculation procedure for estimating the required lime dose and the quantity of sludge produced in tertiary treatment of wastewater with lime is presented. On-site testing for obtaining this information is recommended in specific plant designs.

11.3.2.2 Lime Dose

The extent of phosphate removal is a function of a number of variables including chemical composition of the wastewater and the nature of the phosphate compounds contained. The most important variable controlling phosphate removal and quality of effluent is pH rather than quantity of lime added. Figure 11-1, reference (22), shows a correlation of P-removal as a function of pH, utilizing data from several sources.

Another factor of importance is the hardness level of the wastewater. In some cases (23) it was necessary to go to pH in excess of 11 to get good clarification with the low hardness water. Others (24) obtained satisfactory clarification at pH of about 10 in the hard water.

The chemistry of lime-softening of natural waters is sufficiently well developed that the charge of lime required to achieve a given removal of calcium and magnesium can be predicted and the final pH can be closely estimated. With wastewaters, suspended and dissolved organic materials as well as ortho and polyphosphates apparently affect lime dosage required and can cause super-saturation of precipitating species. Typically, lime dose required to reach a given pH is higher than predicted from softening equations. Likewise, calcium and magnesium content of the product water is higher than would be predicted for lime treated natural waters.

A correlation for estimating lime dose needed to bring a secondary effluent to a given pH is presented in Figure 11-2. It is based on data from several sources (23-27). The pH is reasonably well correlated with the ratio of lime dose to initial alkalinity. Other factors which affect the result are ratio of magnesium to calcium, ionic strength, and the phosphate level. The effects caused by these variables can be estimated, and computer programs have been developed to make the necessary calculation (22) (28). However, the accuracy of these calculations is provisional because of the uncertain effects of contaminants in wastewater.

Figure 11-2 can be utilized to predict lime dose if an estimate of the desired pH level can be made. The curve is entered at the ordinate, and the abscissa, the ratio of CaO dose to alkalinity, is read from the graph. If alkalinity of the water is known, active CaO dose can be calculated. Total quicklime dose can be calculated from active CaO dose and the active CaO content of the quicklime.

11.3.2.3 Sludge Production

The principal reactions for estimating lime sludge production are:

FIGURE 11-1

EFFECT OF pH ON TOTAL RESIDUAL PHOSPHORUS

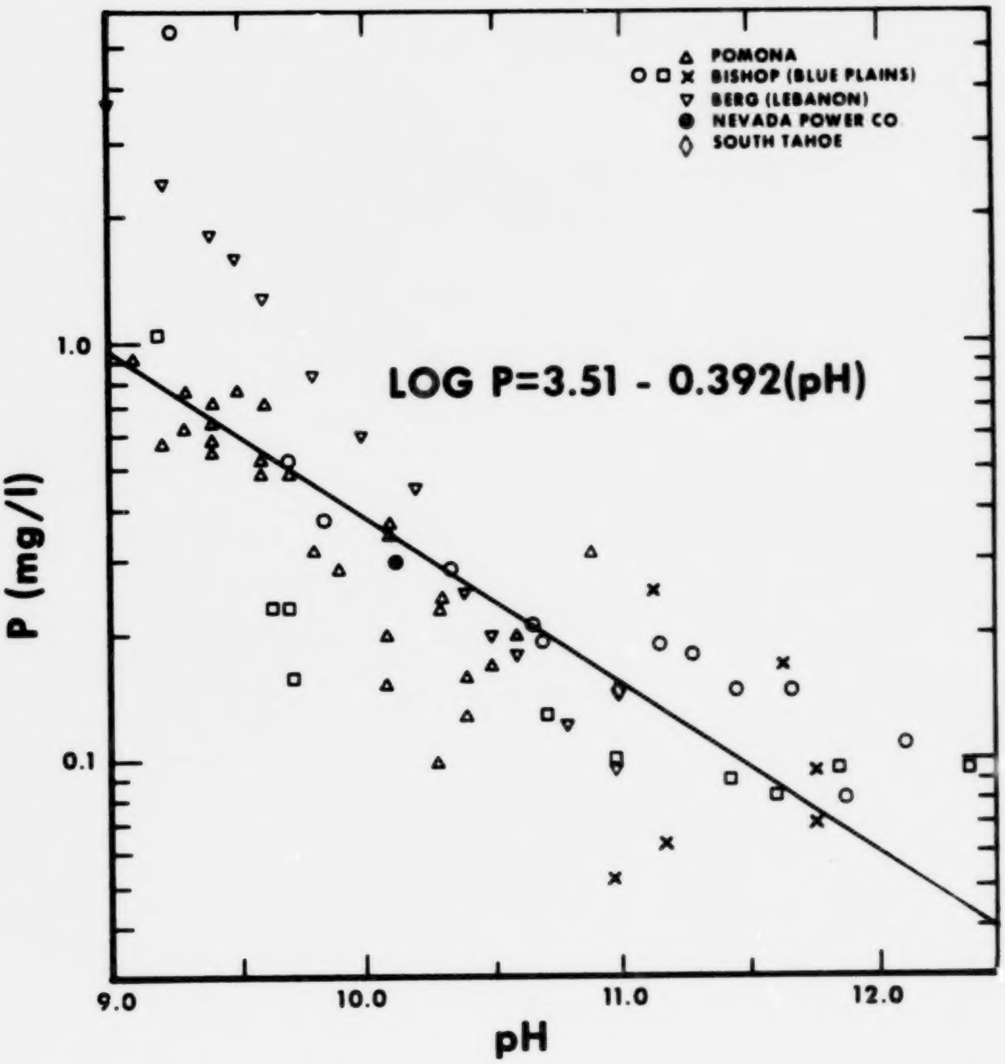
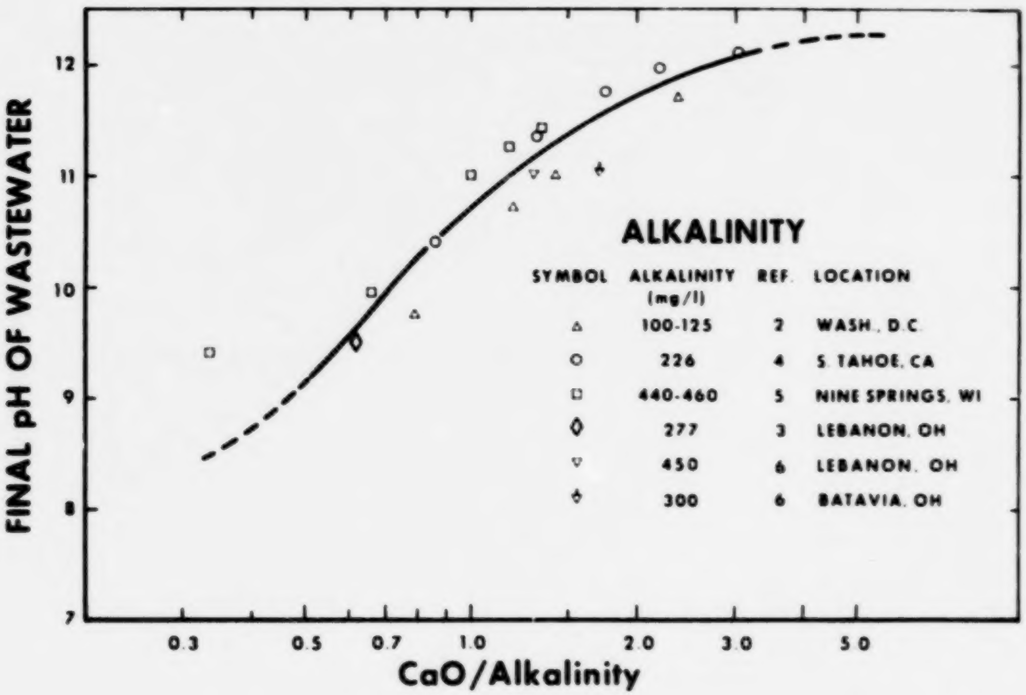


FIGURE 11-2

RATIO OF LIME DOSE TO ALKALINITY (BOTH MG/L)
FOR A PRESCRIBED FINAL pH



$$5Ca(OH)_2 + 3HPO_4^{--} \longrightarrow Ca_5OH(PO_4)_3 + 6OH^- + 3H_2O \quad (1)$$

$$Mg^{++} + 2OH^- \longrightarrow Mg(OH)_2 \quad (2)$$

$$Ca(OH)_2 + HCO_3^- \longrightarrow CaCO_3 + H_2O + OH^- \quad (3)$$

The information required and the steps to take in estimating the sludge production are listed below. Informational input needs are marked with an asterik.

TABLE 11-7

INFORMATION NEEDED TO CALCULATE LIME SLUDGE MASS

Information Needed	Determination
A. P^*_{in}, desired P removal* (sometimes, BOD_{in}, desired BOD removal). Results of jar test or experience*	pH level
B. Alkalinity of wastewater*, pH level, Figure 11-2	CaO dose
C. P_{in}, P removal, Equation 1	$Ca_5OH\ (PO_4)_3$(abbrev.:HAP) precipitated
D. Mg_{in}, from analysis or estimate*, Mg_{out}, from analysis or Figure 11-4, Equation 2	$Mg(OH)_2$ precipitated
E. Ca_{in}, from analysis or estimate*, Ca_{out}, from analysis or Figure 11-3, Ca from lime dose (see item B), Ca removal in HAP (see item C), Equation 3	$CaCO_3$ precipitated
F. pH (see item A)	Verify that sufficient CO_3 is present. If not, estimates of Ca_{out} were in error.

G. Active CaO in lime*	Inerts brought in with quicklime.

The procedures for steps A and B have already been presented. Step C requires no explanation. If estimates of Mg_{out} and Ca_{out} are needed for steps D and E, available data (24) can be used as shown in Figure 11-3 and Figure 11-4. They show the calcium and magnesium levels obtained at the clarifier outlet in long-term evaluation of tertiary lime treatment of secondary effluent. Magnesium (Figure 11-4) precipitation commences at pH 10 and is essentially complete at pH 11. The percent in solution could be estimated as declining from 100% at pH 10 to 0 at pH 11.

The calcium concentration (Figure 11-3) is seen to decline precipitously when pH is increased to 9.5. Concentration then increases as pH is increased further. This behavior is rational in light of knowledge of bicarbonate-carbonate equilibria. When pH is increased, bicarbonate ion is converted to carbonate ion. When the solubility product of $CaCO_3$ is exceeded, precipitation occurs and Ca concentration falls. As pH is increased further by more $Ca(OH)_2$ addition, eventually all of CO_3 is removed by $CaCO_3$ precipitation. Calcium concentration then increases as calcium hydroxide dissolves to the limit of its solubility.

Step E outlines the calculation of calcium carbonate precipitation. The calcium carbonate deposited is calculated from a calcium balance. After considering the amount of calcium removed in the hydroxyapatite ($Ca_5OH\ (PO_2)_3$), the $CaCO_3$ is calculated from the net loss of calcium.

A verification step is advisable to determine whether there is sufficient becarbonate to account for the $CaCO_3$ produced. Bicarbonate can be determined from alkalinity. At pH below about 8.3, all of alkalinity measured is due to bicarbonate. At pH below 7, there is more potential carbonate available due to the presence of dissolved carbon dioxide, which is not measured in the alkalinity test. Estimates of dissolved CO_2 can be made by following procedures outlined in "Standard Methods" (29). If the bicarbonate is insufficient to balance with lost calcium calculated from the material balance, calcium in the final effluent was estimated at too low a value. There is little likelihood of this problem at pH levels below about 10.8.

The entire procedure described, including the verification step and the totaling of all of the components of the sludge, is shown in the illustration in Table 11-8. It is evident from the total quantities listed at the end of the table that calcium carbonate makes up about three-quarters of the mass of the sludge. For this reason, substantial errors could occur in estimates of the other sludge quantities without introducing much error into the final result.

FIGURE 11-3

PRECIPITATION OF CALCIUM BY LIME ADDITION

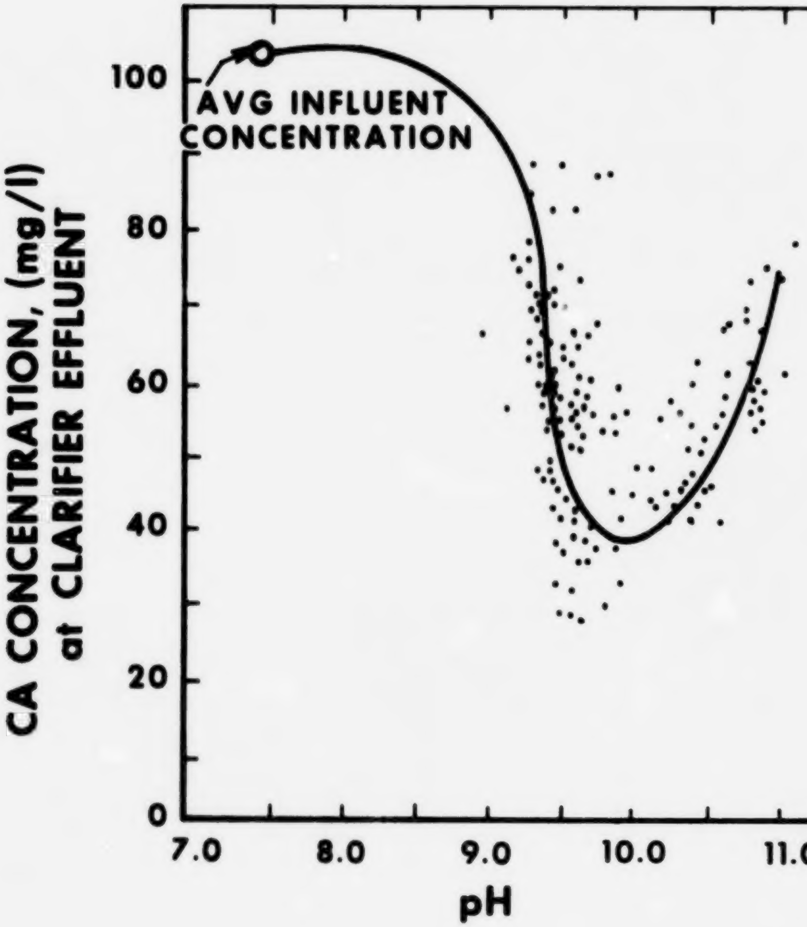


FIGURE 11-4

PRECIPITATION OF MAGNESIUM BY LIME ADDITION

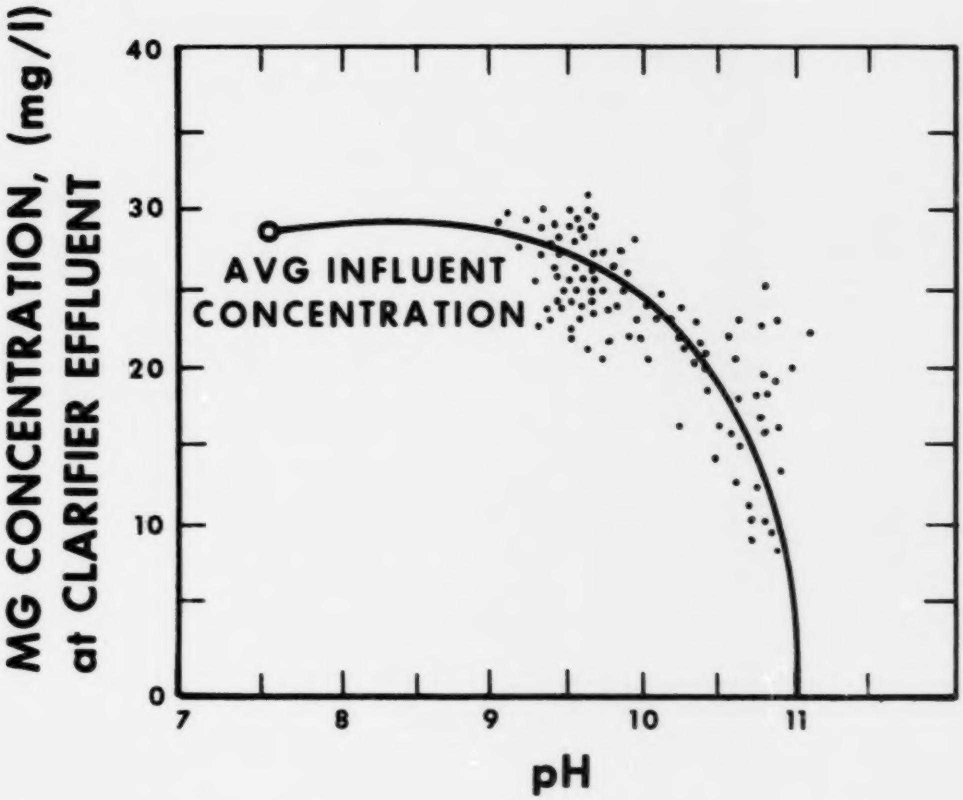


TABLE 11-8

CALCULATION OF LIME SLUDGE MASS

Illustration of Procedure for Estimating Sludge Quantity

Secondary Effluent Characteristics

Phosphate Content	10 mg/l as P
Alkalinity	300 mg/l as $CaCO_3$
Calcium	100 mg/l
Magnesium	20 mg/l

Treatment Conditions

Remove 85% of phosphate by raising pH to 10.5

Final Phosphorus Content	1.5 mg/l as P
Estimate of Ca (Figure 11-3)	50 mg/l
Estimate of Mg (Figure 11-4)	10 mg/l

Lime Dose

From Figure 11-1, 0.92 mg CaO/mg alkalinity at pH 10.5.

Dose = 0.92 x 300 = 276 mg/l CaO

Assume quicklime is 90% active CaO

Total quicklime - 176/0.90 = 307 mg/l

Inerts with quicklime = (0.10)(307) = 30.7 mg/l

Sludge from Phosphate

(10 - 1.5)(502/93) = 45.8 mg/l $Ca_5(OH)(PO_4)_3$

$Mg(OH)_2$ in Sludge

(20 - 10)(58.3/24.3) = 24.0 mg/l $Mg(OH)_2$

Sludge from $CaCO_3$ Precipitation

Input Ca = 100 from sec. effluent + (276)(40/56) from lime dose

Output Ca = 45.8 x (5)(40)/(502) + 50 + Y

(removed with hydroxyapatite)	(remaining in solution)	(precipitated as $CaCO_3$)

Y = 100 + 197 - 18.3 - 50 = 129 mg/l

129 x (100/40) = 323 mg/l $CaCO_3$ precipitated

Total Sludge

Inerts in quicklime	30.7
$Ca_5OH(PO_4)_3$	45.8
$Mg(OH)_2$	24.0
$CaCO_3$	323.0
Total =	433.5 mg/l

Verification that Sufficient CO_3 is Present

300 mg/l alkalinity (as $CaCO_3$)

300 x (1/50) = 6.0 g-equivalents (or g-mols) HCO_3^-

$$HCO_3^- + Ca^{++} + OH^- \longrightarrow CaCO_3 + H_2O$$

6.0 g-mols HCO_3^- can produce 6.0 g-mols $CaCO_3$

6.0 x 100 = 600 mg/l $CaCO_3$

The CO_3 available could produce as much as 600 mg/l $CaCO_3$. This is substantially greater than the amount calculated above (323 mg/l) from the calcium balance. Thus, there is ample CO_3 present. If there were insufficient CO_3 (which can happen at ph of 11 and greater), it would indicate that the estimate of Ca^{++} after treatment was too low. The amount of $CaCO_3$ indicated by the calculation from the alkalinity would be correct.

Additional Sludge from Recarbonation

Recarbonation to pH 9.5 will reduce Ca concentration to about 30 mg/l (see Figure 11-3). For this example, the additional calcium carbonate sludge produced is calculated as follows:

$$(50 - 30)(100/40) = 50 \text{ mg/l } CaCO_3$$

11.4 Dewatering at Specific Locations

11.4.1 Tertiary Treatment with Lime

11.4.1.1 South Tahoe

At South Tahoe, California, wastewater is treated in conventional primary, followed by a waste activated sludge plant. Phosphate is removed by tertiary lime treatment to pH 11 with recarbonation. A comprehensive report describing the performance of the Tahoe plant and sludge dewatering aspects has been published (25).

The tertiary lime sludge is dewatered and classified at Lake Tahoe by means of solid bowl centrifuges. Sludge is thickened to 8 percent solids. It then flows at about 20 gpm to a 75 HP Bird 24" x 60" solid-bowl concurrent-flow centrifuge. The centrifuge is deliberately operated with high losses to the centrate in order to reject magnesium hydroxide, hydroxyapatite and organic solids, leaving behind a high concentration of calcium carbonate in the cake. The cake is calcined and is recovered. The centrate is sent to a second centrifuge, polymer is added, and solids are removed as a 30% solids cake.

The fact that a centrifuge tends to reject fine low-density solids may lead to fears that fine solids will build up in the system. This has apparently not occured at South Tahoe.

11.4.1.2 Blue Plains Pilot Plant

At the Blue Plains pilot plant, a substantial amount of information was obtained on the sludges produced by treating secondary effluent with lime. This information will be published soon as a report in the EPA series (30).

Secondary effluent from the District of Columbia's Blue Plains Treatment Plant was treated with lime to a pH range of 11 to 12 for phosphate removal. Instead of utilizing second-stage recarbonation to remove excess Ca ions and recover additional calcium carbonate, sodium carbonate was added to the single clarifier for this purpose. The sludge produced was continuously thickened, vacuum filtered, calcined, slaked and recycled to the tertiary clarifier to supply the bulk of the lime demand. Information on sludge thickening and dewatering was obtained with and without recycle of reclaimed lime.

Most of the data during recycle operations were obtained at 56 percent recovery; that is, 56 percent of the active lime dose consisted of recycled lime. Recycle was approximately 3 pounds of recycled solids to one pound of fresh lime. Lime dose consisted of 300 mg/l of active lime plus 240 mg/l of inerts.

Buildup of inerts did not affect phosphorus removal but increased solids loading through the sludge handling system. Solids content of sludge removed from the clarifier was 20 g/l for no-recycle operation and 32 g/l for 56 percent recycle.

The sludge was thickened in a 6-foot diameter vessel equipped with pickets. Sludge thickened to 15-20% solids for recycle as well as no-cycle operations, despite solids loading rates over 70 percent higher for the recycle case.

Information on filtration was obtained from filter leaf tests. Equations for form rate for the recycle and no-recycle cases were developed. These equations, rearranged slightly to give results in terms of yield and cycle time, are:

No Recycle: $Y = (2.18)(f)^{0.5}(C)^{0.84}(p)^{0.21}/(\phi)^{0.5}$

Recycle: $Y = (3.28)(f)^{0.5}(C)^{0.80}(p)^{0.26}/(\phi)^{0.5}$

Where Y = yield, lb/hr-sq ft

ϕ = cycle time (min), or reciprocal of rpm

C = Solids concentration, wt %

p = operating vacuum, in. Hg

f = fraction submergence

Calculating filter yield at the following conditions,

$f = 0.25$

C = 15% solids

p = 15" Hg vacuum

ϕ = 1 min. (i.e., 1 rpm)

gives the following result:

No Recycle: Y = 18.7 lb/hr-sq ft

Recycle: Y = 29.0 lb/hr-sq ft

This study reports substantial cake cracking which reduced vacuum across the filter. Nevertheless, solids contents from 30-40% were generally obtained. It was also noted that, after a process upset which caused biological solids to carry over into the tertiary treatment system, filter yield deteriorated drastically. However, filtration was possible without chemical conditioning.

11.4.1.3 Butler, Indiana

At Butler, wastewater is treated in a 0.4 mgd conventional primary plus trickling filter plant. Primary settling precedes two-stage biological filters. Conventional sludge is digested and dried on beds. Phosphorus is removed by tertiary lime in a reactor-clarifier at pH 10. Chemical dose is 150 mg/l quicklime plus ferric chloride and polymer to aid settling. Total phosphorus removal is about 80 percent. Sludge is not dewatered and is pumped to a lagoon.

The Butler plant is somewhat unique in that the wastewater is quite hard (300 mg/l) and its magnesium content is high (30 mg/l). Ordinarily, good phosphorus removal is achieved in water of this type at a lower pH than in soft low-magnesium water. Required phosphorus removal was only 80% and a pH of 10 was sufficient to get satisfactory performance.

Sludges leaving the reactor-clarifier were about 8.4 percent solids. Overnight settling increased this to 9.0 or 9.5 percent. Lagoon sludge, even after only a few days settling, was about 17% suspended solids. Fresh sludge from the reactor-clarifier could be filtered at 5.6 lb/hr-sq ft at a cycle time of 2.9 min and 33% immersion. Higher yield could be achieved at lower cycle times but this produced cakes less than one-eighth inch thick, which caused problems with cake release.

Sludge was mostly mineral in nature. Percent volatile solids was under 10%. Magnesium content of the sludge was low. At higher pH, more magnesium would have been precipitated, which could have produced a poorer filtering sludge.

11.4.2 Lime to Primary Settler

11.4.2.1 Contra Costa

The Contra Costa Sanitary District, Walnut Creek, California, has carried out pilot and plant scale evaluations of an integrated process for handling sludges produced by adding lime to primary wastewater treatment. This work has been reported (31) and is discussed in Chapter 5.

The treatment process comprises lime flocculation with polymer or ferric chloride followed by primary sedimentation, activated sludge under conditions such that nitrification is achieved, denitrification with methane addition, and chlorination. Waste activated sludge is recycled to the raw sewage to be removed from the primary clarifier along with the chemical-primary sludge. Denitrification sludge is also returned to the incoming raw sewage, but its amount is negligible.

In addition to investigations of process performance, studies were made to evaluate (a) predictive methods for estimating sludge production, (b) the degree of thickening achieved in primary clarification, (c) the utility of two-step procedure in which sludge is centrifugally classified to recover calcium carbonate and the centrate is centrifuged again to remove solids, and (d) anaerobic digestion of the first-stage centrate.

Sludge mass predicted from analysis of influent and effluent analyses checked satisfactorily with measured sludge mass. Operation at pH 10.2 produced a clarifier sludge of about 4.0 percent solids. At pH 11, higher sludge solids could be attained. Centrifugal classification achieved an 80% recovery of calcium carbonate in relatively dry cakes of 42-57% solids. Solids remaining in the centrate could be removed by polymer-assisted second-stage centrifugation at recoveries as high as 90%. Results were not as good at pH greater than 11. The centrates from the first centrifugation could be anaerobically digested, although volatile solids reduction was lower than for conventional sludge processing. In addition, precipitated magnesium redissolved in digester supernatant. Return of supernatant to process would increase lime demand and magnesium would be precipitated again. It is believed that handling of supernatant by separate treatment or return to the secondary process would avoid that problem.

11.4.2.2 Eimco at Salt Lake City

Eimco Division of Envirotech has conducted a 100 gpm pilot plant investigation of a wastewater treatment system in which chemicals are added in the primary step and soluble organic material is removed by powdered activated carbon. A progress report on this investigation has recently been published (14).

Comparisons of sludge mass calculated from clarifier input and output analyses with measured mass showed reasonable agreement. When iron was added to the primary, measured mass was 20% higher than the calculated value. However, this additional mass appeared to include removal of soluble organics. When adding alum, calculated value (including soluble organic material) was virtually the same as the measured value. For lime, measured values averaged 20% higher than calculated values.

Thickening and filtration data were incomplete for iron and alum. However, data with lime were developed. After 24 hours thickening, sludges with pH values greater than 11.5 thickened to 2.0 to 9.8% solids, with median near 7%. For pH values less than 11.0, sludge thickened to 9.0 to 18.0% solids, with median near 14%.

Filtration yields were obtained for high and low pH operation. No polymer was needed. For low pH sludge at the 14% thickened solids content, maximum filter yield was 6.0 lb/hr-sq ft. For high pH sludge at the 7% thickened solids content, maximum filter yield was 3.0 lb/hr-sq ft.

It should be noted that Salt Lake City wastewater is high in magnesium. Operation at pH greater than 11.5 precipitates much more magnesium than precipitated at pH less than 11.0. Since magnesium hydroxide is a flocculant precipitate, this could account for the low solids content of thickened sludge from high pH operation.

11.4.2.3 Cleveland Westerly

Extensive pilot plant work was conducted at Cleveland's Westerly plant preliminary to preparing a design for a physical-chemical treatment plant. A report describing this work is in progress (32).

Most of the work was done with lime added prior to the primary to reach a pH of 10.5. Sludge was dewatered using a Bird pilot plant centrifuge and an Eimco belt-type pilot plant vacuum filter.

Sludge solids content leaving the clarifier was in the range of 6% to 14% solids. Thickening prior to centrifugation was not needed. Solids recovery in the centrifuge was poor without polymer. However, polymer doses of less than 2 lb/ton gave recoveries of 95%. Solids content of the sludge cake ranged from 20 to 28 percent.

The lime-primary sludge could be filtered without polymers but performance was improved greatly by addition of small quantities of polymer. Filtration yields in excess of 10 lb/hr-sq ft were recorded. Recoveries of solids were in the order of 95%. Cake solids ranged from 20 to 35% with 28% a typical value.

11.4.2.4 Canadian Experience

The Province of Ontario's Ministry of the Environment has carried out a number of plant-scale testing programs at sewage treatment plants. This work has been summarized in several reports (33-35). Some of that information (34) is summarized here.

The use of lime in activated sludge plants increased sludge production from a normal level of about 180 mg/l to 460 mg/l during addition of 200 mg/l slaked lime. Sludge solids content increased from 4.5% to 9%. In primary plants, sludge production increased from 80 mg/l to 240 mg/l. Sludge solids content increased from 7% to 15%.

Some difficulty was encountered with anaerobic digestion because of erratic dosage with high pH sludge. Subsequent studies have shown that holding the sludge longer in the clarifier results in a fall in sludge pH. Digester operation is then normal except for high solids content of the sludge.

Raw lime sludge was dewatered successfully in a continuous solids bowl conveyor-type centrifuge at the Newmarket/East Gwillimbury plant (36). A detailed investigation showed high recoveries could be achieved with polymer addition. At these high recoveries, phosphorus removal from the centrate was also high (37).

With lime added to the primary of a conventional plant (primary plus activated sludge), filter yield increased from 5.2 to 7.2 lb/hr-sq ft, and conditioning costs (lime plus $FeCl_3$) fell from \$16/ton down to \$11/ton. Cake solids content was 29%.

11.4.3 Alum and Iron Salts

11.4.3.1 Blue Plains

Alum was added to the 300 mgd wastewater treatment plant at Blue Plains, Washington, D. C., for several months. At this plant, primary treatment is followed by high rate activated sludge. Waste activated sludge is combined with sludge withdrawn from the clarifier, and the mixed sludge is thickened and digested. The digested sludge was elutriated in two stages using polymer to retain solids. The sludge was filtered using polymer and ferric chloride for conditioning. The following information on alum addition tests is based on discussions with plant operating personnel.

Alum was added to the aerator in gradually increasing dose until the average dose was 40 mg/l of alum (9.1% Al). This developed a relatively high level of aluminum in the sludge because the Blue Plains wastewater is weak.

Alum addition did not affect thickening. Sludge leaving the thickeners remained at 7.5% solids. Digestion was likewise unaffected. Suspended solids leaving the digesters remained at 3.5%. Elutriation could be carried out but a high polymer dose was required to retain suspended solids. Chemical costs for elutriation increased from \$3/ton before alum addition to \$7/ton during the test. Filter yields were about the same as with prior operation, but solids content of the cake dropped from 23% to 20%. Polymer condition cost for filtration, which had been \$6/ton, rose to \$7/ton.

11.4.3.2 Pomona

At Pomona, California, a physical-chemical plant has been in operation to produce a feed stream for a long-term test of carbon columns. The wastewater has been treated in a primary clarifier with 22 mg aluminum per liter. A field test was conducted to determine the dewatering characteristics of this sludge. That unpublished report (38) is summarized here.

Gravity thickening tests showed that, without polymer, sludge settled slowly. Thickening rate to produce a 4% solids sludge was quite low at 6.6 lb/day-sq ft. With polymer addition, that rate increased to 20 lb/day-sq ft with 2 lb/ton of anionic polymer and to 70 lb/day-sq ft with 6 lb/ton. Filtration rates with 12% lime and polymer were less than 2 lb/hr-sq ft at the low cycle time of 2 min. However, the sludge employed was only 2% solids. Rate would be expected to increase proportionately if thicker sludge were available.

One of the reasons this field test was conducted was to verify other data indicating that sludges produced when metals are added to the primary settle poorly and have lower filter yields than conventional sludge. The Pomona field test corroborated these findings.

11.4.3.3 Grand Rapids

The addition of ferric chloride at various points in the process was studied at Grand Rapids, Michigan. The Grand Rapids plant has a conventional primary followed by activated sludge. The ferric chloride was added in three different modes: to the primary alone, split between the primary and the aerator, and to the aerator alone. An investigation of sludge dewatering was not part of this project. However, it was hoped that the filter yield data routinely collected at this plant could be related back to the chemical dosing conditions.

Monthly average filter yields, cake solids, and chemical costs were supplied by Grand Rapids staff and are presented in Figure 11-5. Large fluctuations in all three filtration variables are evident; however, they are not correlated with ferric chloride dosing. This finding was predicted by Grand Rapids staff who observed that the basic nature of the sludge can change from month to month and that necessary maintenance can cause step changes in performance. Some of the interferences noted at Grand Rapids were:

a. Periodic dumping of lime sludge from a water plant into the wastewater collection system.

b. Performance changes in digestion.

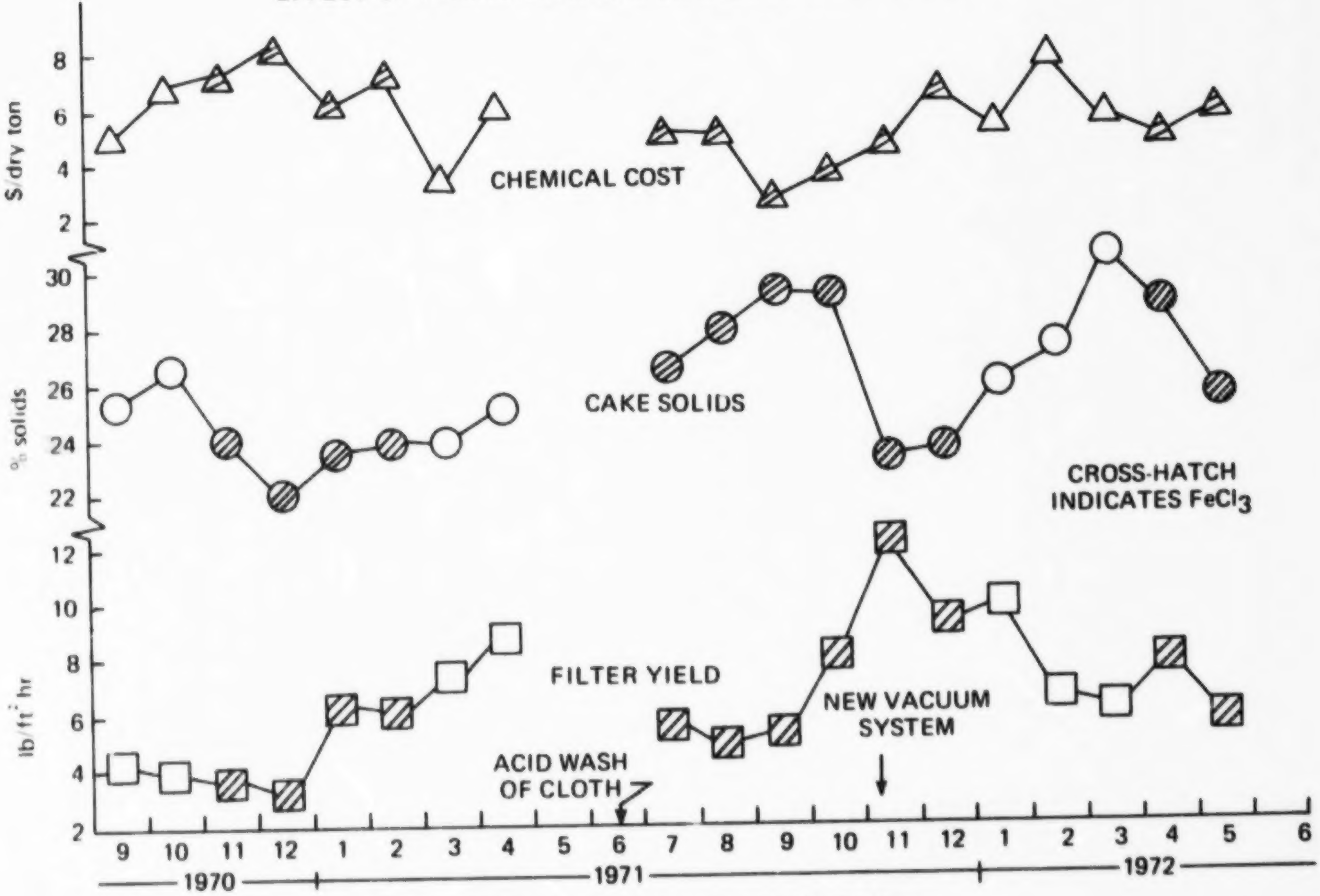
c. Acid cleaning of filter cloths.

d. Cleaning of interior piping of vacuum filters.

e. Replacement of vacuum system.

FIGURE 11-5

EFFECT OF FERRIC CHLORIDE ON SLUDGE DEWATERING



EFFECT OF $FeCl_3$ ADDITION AT GRAND RAPIDS

f. Periodic switch-over from the incinerators, which had a lower capacity than the vacuum filters and needed a dry cake, to landfill, where maximum output was desired and a dry cake was not so important.

11.4.3.4 Canadian Experience

Reference has been made in Section 11.4.2.4 to the work reported by Ontario's Ministry of the Environment. Additional information from a summary paper (34) is included here.

When metal salts were utilized in activated sludge secondary plants, the dry solids sludge production increased 5% to 25%. Sludge concentration dropped slightly to 3.5 to 5.0% from the normal 4.0 to 5.5%.

In primary plants, sludge production increases substantially because of increased solids capture. An increase of 100% was typical and 20% reduction in sludge solids concentration was reported.

In all cases where metal salts were used to precipitate phosphorus, normal digestion was observed. Problems which developed were the result of increased hydraulic and volatile solids loading.

Filtering rates were approximately halved for a primary plant when alum was added. Lime and ferric chloride conditioning costs increased from $3.10/ton to $9.50/ton. Solids content of cake dropped from 31% down to 19%. For a secondary plant with alum added to the primary, filtering rate fell from 5.2 to 4.8 lb/hr-sq ft, and conditioning chemical cost increased slightly. Cake solids was typically 15.9% with alum addition.

11.4.3.5 Chapel Hill

Chapel Hill has a 5 MGD trickling filter plant which is conveniently split into two parallel treatment plants, each with its own primary clarifier, high rate trickling filter, and final clarifier. During an experimental program, the wastewater from one train was dosed with alum for phosphorus removal, just before it entered the final clarifier. A field trip was arranged to this unique facility to determine the impact of alum addition on the wastewater sludge. Unpublished results (39) are summarized here.

Data obtained indicated sludge from the secondary clarifier following the alum-dosed train was essentially unfilterable, even with polymer addition. Since this was a high-rate filter with substantial recirculation before the clarifier, the sludge was very low in organic matter. The control secondary clarifier sludge showed reasonable filter yields. Secondary sludges from both trains were recirculated to their respective primary clarifiers and removed as mixed primary-secondary sludges. Filter leaf tests showed that there was no substantial difference in the filter yields from the two trains.

Data supplied by University of North Carolina staff who operate the plant indicated that 4.75 lb extra sludge was generated per lb of Al added. They also indicated that difficulty was encountered during alum trials in their two-stage digester: the solids content of the supernatant leaving the second digester was unusually high.

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APPENDIX A

EVALUATING PHOSPHORUS REMOVAL STRATEGIES

A computer model for evaluating a number of strategies for removing phosphorus in wastewater has been developed for EPA.[1] This model reports to the user the total cost of a selected strategy for removing phosphorus. 22 treatment schemes can be selected and evaluated depending on local conditions. These schemes are:

1. Alum & Polymer Addition to Primary Settler
2. Ferric Chloride & Polymer Addition to Primary Settler
3. Ferrous Chloride & Lime Addition to Primary Settler
4. Lime Addition to Primary Settler
5. Alum & Polymer Addition to Flocculation Basin
6. Ferric Chloride & Polymer Addition to Flocculation Basin
7. Ferrous Chloride & Lime Addition to Flocculation Basin
8. Lime Addition to Flocculation Basin
9. Alum Addition to Aeration Basin
10. Ferric Chloride Addition to Aeration Basin
11. Sodium Aluminate Addition to Aeration Basin
12. Alum Addition to Aeration Basin & Multi-Media Filtration
13. Ferric Chloride Addition to Aeration Basin & Multi-Media Filtration
14. Sodium Aluminate Addition to Aeration Basin & Multi-Media Filtration
15. Alum Addition After Trickling Filter
16. Ferric Chloride Addition After Trickling Filter
17. Alum Addition After Trickling Filter & Multi-Media Filtration
18. Ferric Chloride Addition After Trickling Filter & Multi-Media Filtration
19. Alum Addition to Flocculation Basin After Conventional Secondary Treatment
20. Ferric Chloride Addition to Flocculation After Conventional Secondary Treatment
21. Lime (1-Stage) Addition to Flocculation After Conventional Secondary Treatment
22. Lime (2-Stage), Addition to Flocculation Basin After Conventional Secondary Treatment

Four levels of effluent phosphorus concentration (total unfiltered phosphorus) were assumed to be achievable depending upon the type of treatment provided. These are:

2.0 mg/l (Option 1) Chemical addition to the primary clarifier. (Treatment schemes 1-4)

(Option 2) Chemical addition to a flocculation basin prior to the primary clarifier. (Treatment schemes 5-8)

(Option 3) Chemical addition to the aeration basin. (Treatment schemes 9-11)

---

[1] "A computer Model for Evaluating Community Phosphorous Removal Strategies," prepared for U.S. EPA by JBF Scientific Corp., Burlington, Mass., Report No. 400/9-73/001 (Oct. 1973).

(Option 4) Chemical addition after the trickling filter. (Treatment schemes 15-16)

0.5 mg/l (Option 5) Chemical addition to the aeration basin plus multi-media filtration. (Treatment schemes 12-14)

or

Chemical addition after the trickling filter plus multi-media filtration. (Treatment schemes 17-18)

(Option 6) Addition of lime to a flocculation basin following conventional secondary treatment. (Treatment scheme 21)

0.3 mg/l (Option 7) Addition of alum or ferric chloride to a flocculation basin following conventional secondary treatment—one stage. (Treatment schemes 19-20)

0.1 mg/l (Option 6) Addition of lime to a flocculation basin following conventional secondary treatment—two stages. (Treatment scheme 22)

The model has been designed specifically for planning purposes and can be used by municipalities, planning agencies and government policy organizations. In order to insure its usefulness, EPA plans to periodically update the model as additional performance data on phosphorus removal processes becomes available.

The phosphorus removal model (REMOVE) is designed to provide an estimate of the cost to a particular community of achieving a designated level of phosphorus removal. The program begins with the entry of data describing the community in terms of population and future trends, the treatment plant in terms of present size and excess capacity, and the influent wastewater characteristics. The data can be entered interactively through a subroutine (SYSTEM).

After input data have been entered, the user has the opportunity to investigate various control strategies for achieving a low phosphorus concentration in the wastewater effluent. Typically, the effects on cost of reducing phosphorus in detergent products, of increasing or decreasing the amount of government financing available, or of using various treatment options can be investigated.

It is important to note that the model has been developed as a simulation only and as such does not achieve a mathematical optimization.

From the description of the community, treatment plant, wastewater characteristics, and desired strategy the model will design a treatment plant incorporating the present plant as described by the user and will add those processes or additional equipment necessary to achieve the desired effluent level. The program converts the sizing information into construction costs, operating man-hours, maintenance man-hours, and total material and supply costs for each process.

The following pages illustrate the type of print-out obtained when using the computer model. The particular cases illustrated here are a 5 mgd trickling filter plant and a 20 mgd activated sludge treatment plant.

```
                    5 MGD TRICKLING FILTER PLANT
                   TREATMENT COSTS IN CENTS/1000 GAL.

P                  2       2        2        0.5      0.5      0.5
GOVFF                      .8       .8                 .8       .8
RESTRICT                            yes                         yes

SCHNO
1               13.9    11.9      8.4
2               10.0     8.4      6.6
3               17.8    15.0     13.7
4               13.1    10.9     11.7
5               14.8    12.3      8.8
6               11.0     8.9      7.0
7               18.8    15.3     14.0
8               14.0    11.2     12.1
9               12.3    10.6      7.5
10              10.0     8.4      6.3
11              15.1    13.5      8.8
12                                         18.9     14.3     10.2
13                                         16.0     12.1      9.1
14                                         21.7     17.1     11.5
15               7.7     6.7      4.6
16               8.2     7.5      4.8
17                                         13.7     10.0      7.3
18                                         14.9     11.2      7.7
21                                         15.3     10.3     11.2

NOTES:
1. P = Phosphorus effluent level in mg/l.
2. GOVFF = Fraction of construction cost financed
           by federal and state governments.
3. RESTRICT = Restrictions on phosphate detergents.
4. SCHNO = Treatment scheme number.
5. ___ = Lowest treatment cost.
```

```
                                  CASE  1


LIQUID TREATMENT SCHEME =     15
AMORTIZATION RATE =0.05 AMORTIZATION LIFETIME =20.0
FRACTION OF CONSTRUCTION COST FINANCED BY GOVERNMENT  = .0

PROCESS NAME    CONSTRUCTION         SIZING               AMORTIZED COST      YEAR PROCESS
               COST(1000 $ )       PARAMETER              (1000 $ /TP)       FIRST NEEDED

RAW WSTE PMP         0.0             0.0     MGD              0.0              1994.0
PRE TRETMENT         0.0             0.0     MGD              0.0              1994.0
PRIM SETTLER         0.0             0.0     TSF              0.0              1994.0
PRMYSLGE PMP         0.0             0.0     GPM              0.0              1994.0
TRKLNG FILTR         0.0             0.0     TCF              0.0              1994.0
SECY SETTLER         0.0             0.0     TSF              0.0              1994.0
RECIRC PMPNG         0.0             0.0     MGD              0.0              1994.0
CHLORNE FEED         0.0             0.0     #/D              0.0              1994.0
CHLORNE BASN         0.0             0.0     TCF              0.0              1994.0
GRAV THICKNR        34.824           0.022   TSF              2.794            1974.0
SLGE HLD TNK        15.649           0.987   TCF              1.256            1974.0
VACUUM FILTR       104.751          39.600   SQFT             8.406            1974.0
ALUM FEEDING        49.858         624.750   #/D              4.001            1974.0
FECL FEEDING        45.525          64.565   #/D              3.653            1974.0
LIME FEEDING        15.901         746.367   #/D              1.276            1974.0

TOTALS =           266.507                                   21.385
```

```
                                PERIODIC OPERATING COSTS

PROCESS NAME      OPERATING MAN-HOURS     MAINTAINANCE MAN-HOURS
                   (MAN-HOURS/TP)            (MAN-HOURS/TP)
                 AVE.     MAX.     MIN.     AVE.     MAX.     MIN.

RAW WSTE PMP       0.       0.       0.       0.       0.       0.
PRE TRETMENT       0.       0.       0.       0.       0.       0.
PRIM SETTLER       0.       0.       0.       0.       0.       0.
PRMYSLGE PMP       0.       0.       0.       0.       0.       0.
TRKLNG FILTR       0.       0.       0.       0.       0.       0.
SECY SETTLER       0.       0.       0.       0.       0.       0.
RECIRC PMPNG       0.       0.       0.       0.       0.       0.
CHLORNE FEED       0.       0.       0.       0.       0.       0.
CHLORNE BASN       0.       0.       0.       0.       0.       0.
GRAV THICKNR       2.       2.       2.       1.       1.       1.
SLGE HLD TNK      18.      18.      18.      11.      11.      11.
VACUUM FILTR     665.     665.     665.      80.      80.      80.
ALUM FEEDING       0.       0.       0.    1473.    1473.    1473.
FECL FEEDING       0.       0.       0.     600.     600.     600.
LIME FEEDING       0.       0.       0.    1578.    1578.    1578.

TOTALS =         685.     685.     685.    3742.    3742.    3742.
```

PERIODIC OPERATING COSTS

PROCESS NAME	MATERIAL & SUPPLY COST ($ /TP) AVE.	MAX.	MIN.	LABOR COST (AVERAGE) (1000$/TP)	TOTAL O&M (1000$/TP)
RAW WSTE PMP	0.0	0.0	0.0	0.0	0.0
PRE TRETMENT	0.0	0.0	0.0	0.0	0.0
PRIM SETTLER	0.0	0.0	0.0	0.0	0.0
PRMYSLGE PMP	0.0	0.0	0.0	0.0	0.0
TRKLNG FILTR	0.0	0.0	0.0	0.0	0.0
SECY SETTLER	0.0	0.0	0.0	0.0	0.0
RECIRC PMPNG	0.0	0.0	0.0	0.0	0.0
CHLORNE FEED	0.0	0.0	0.0	0.0	0.0
CHLORNE BASN	0.0	0.0	0.0	0.0	0.0
GRAV THICKNR	5.67	5.67	5.67	0.013	0.019
SLGE HLD TNK	54.57	54.57	54.57	0.133	0.187
VACUUM FLITR	3166.75	3166.75	3166.75	3.382	6.549
ALUM FEEDING	78872.44	78872.69	78872.69	6.635	85.505
FECL FEEDING	2998.12	2998.12	2998.12	2.725	5.724
LIME FEEDING	2855.35	2855.35	2855.35	7.167	10.022
TOTALS =	87952.75	87953.00	87953.00	20.109	106.062

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```
                         UNIT COST DATA
                      (CENTS/1000 GALLONS)

PROCESS NAME   AMORTIZATION   LABOR   MATERIAL + SUPPLY   TOTAL

RAW WSTE PMP       0.0         0.0          0.0            0.0
PRE TRETMENT       0.0         0.0          0.0            0.0
PRIM SETTLER       0.0         0.0          0.0            0.0
PRMYSLGE PMP       0.0         0.0          0.0            0.0
TRKLNG FILTR       0.0         0.0          0.0            0.0
SECY SETTLER       0.0         0.0          0.0            0.0
RECIRC PMPNG       0.0         0.0          0.0            0.0
CHLORNE FEED       0.0         0.0          0.0            0.0
CHLORNE BASN       0.0         0.0          0.0            0.0
GRAV THICKNR       0.15        0.00         0.00           0.15
SLGE HLD TNK       0.07        0.01     [illegible]    [illegible]
VACUUM FILTR       0.46        0.19         0.17       [illegible]
ALUM FEEDING       0.22    [illegible]  [illegible]    [illegible]
FECL FEEDING       0.20        0.15     [illegible]    [illegible]
LIME FEEDING       0.07    [illegible]  [illegible]    [illegible]

TOTALS =           1.17    [illegible]  [illegible]    [illegible]

                        SOLID SINKS COSTS

TIME      SINK    AMOUNT SOLIDS         COST
PERIOD    USED       (TONS)            ($/TP)
   1        1         410.46         10261.50
   2        1         410.46         10261.50
   3        1         410.46         10261.50
   4        1         410.46         10261.50
   5        1         410.46         10261.50
   6        1         410.46         10261.50
   7        1         410.46         10261.50
   8        1         410.46         10261.50
   9        1         410.46         10261.50
  10        1         410.46         10261.50
  11        1         410.46         10261.50
  12        1         410.46         10261.50
  13        1         410.46         10261.50
  14        1         410.46         10261.50
  15        1         410.46         10261.50
  16        1         410.46         10261.50
  17        1         410.46         10261.50
  18        1         410.46         10261.50
  19        1         410.46         10261.50
  20        1         410.46         10261.50

 TOTALS =            8209.17       205229.12

UNIT COSTS FOR SOLID WASTE DISPOSAL (CENTS/1000  GAL.) = 0.50
TOTAL UNIT COST (CENTS/1000 GAL) = 7.66
```

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20 MGD ACTIVATED SLUDGE PLANT
TREATMENT COSTS IN CENTS/1000 GAL.

P	2	2	2	0.5	0.5	0.5
GOVFF		.8	.8		.8	.8
RESTRICT			yes			yes
SCHNO						
1	11.1	9.9	6.4			
2	7.6	6.5	4.8			
3	16.5	13.9	12.7			
4	9.8	8.6	9.5			
5	11.7	10.1	6.7			
6	8.1	6.7	5.1			
7	17.1	14.2	12.9			
8	10.4	8.9	9.7			
9	8.5	8.0	5.0			
10	6.2	5.9	4.0			
11	11.2	11.0	6.4			
12				13.4	10.9	7.0
13				11.3	8.8	6.0
14				16.3	13.8	8.4
15	7.9	6.4	4.4			
16	8.4	7.0	4.7			
17				12.3	8.9	6.3
18				13.6	10.0	6.8
21				9.4	6.2	7.1

NOTES:
1. P = Phosphorus effluent level in mg/l.
2. GOVFF = Fraction of construction cost financed by federal and state governments.
3. RESTRICT = Restrictions on phosphate detergents.
4. SCHNO = Treatment scheme number.
5. ___ = Lowest treatment cost.

CASE 2

```
LIQUID TREATMENT SCHEME =     10
AMORTIZATION RATE =0.05 AMORTIZATION LIFETIME =20.0
FRACTION OF CONSTRUCTION COST FINANCED BY GOVERNMENT  = .0

PROCESS NAME     CONSTRUCTION        SIZING                  AMORTIZED COST      YEAR PROCESS
                 COST(1000 $ )      PARAMETER                 (1000 $ /TP)       FIRST NEEDED

RAW WSTE PMP          0.0              0.0     MGD                0.0               1994.0
PRE TRETMENT          0.0              0.0     MGD                0.0               1994.0
PRIM SETTLER          0.0              0.0     TSF                0.0               1994.0
PRMYSLGE PMP          0.0              0.0     GPM                0.0               1994.0
AERTN BASIN           0.0              0.0     TCF                0.0               1994.0
MECH AERATN           0.0              0.0      HP                0.0               1994.0
SECY SETTLER          0.0              0.0     TSF                0.0               1994.0
RECIRC PMPNG          0.0              0.0     MGD                0.0               1994.0
CHLORNE FEED          0.0              0.0     #/D                0.0               1994.0
CHLORNE BASN          0.0              0.0     TCF                0.0               1994.0
GRAV THICKNR         38.163            0.540   TSF                3.062             1974.0
SLGE HLD TNK          0.0              0.0     TCF                0.0               1994.0
VACUUM FILTR         98.759           34.950   SQFT               7.925             1974.0
FECL FEEDING        115.133         6446.578   #/D                9.239             1974.0
LIME FEEDING         36.171         2541.451   #/D                2.903             1974.0

TOTALS =            288.226                                      23.128
```

```
                                    PERIODIC OPERATING COSTS

PROCESS NAME       OPERATING MAN-HOURS           MAINTAINANCE MAN-HOURS
                     (MAN-HOURS/TP)                  (MAN-HOURS/TP)
                    AVE.      MAX.      MIN.      AVE.      MAX.      MIN.

RAW WSTE PMP          0.        0.        0.        0.        0.        0.
PRE TRETMENT          0.        0.        0.        0.        0.        0.
PRIM SETTLER          0.        0.        0.        0.        0.        0.
PRMYSLGE PMP          0.        0.        0.        0.        0.        0.
AERTN BASIN           0.        0.        0.        0.        0.        0.
MECH AERATN           0.        0.        0.        0.        0.        0.
SECY SETTLER          0.        0.        0.        0.        0.        0.
RECIRC PMPNG          0.        0.        0.        0.        0.        0.
CHLORNE FEED          0.        0.        0.        0.        0.        0.
CHLORNE BASN          0.        0.        0.        0.        0.        0.
GRAV THICKNR         41.       41.       41.       23.       23.       23.
SLGE HLD TNK          0.        0.        0.        0.        0.        0.
VACUUM FILTR       3013.     3013.     3013.      369.      369.      369.
FECL FEEDING          0.        0.        0.     2884.     2884.     2884.
LIME FEEDING          0.        0.        0.     2009.     2009.     2009.

TOTALS =           3055.     3055.     3055.     5284.     5284.     5284.
```

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```
                         PERIODIC OPERATING COSTS

PROCESS NAME        MATERIAL & SUPPLY COST          LABOR COST    TOTAL O&M
                           ( $ /TP)                 (AVERAGE)    (1000$/TP)
                 AVE.        MAX.          MIN.    (1000$/TP)

RAW WSTE PMP       0.0         0.0          0.0         0.0           0.0
PRE TRETMENT       0.0         0.0          0.0         0.0           0.0
PRIM SETTLER       0.0         0.0          0.0         0.0           0.0
PRMYSLGE PMP       0.0         0.0          0.0         0.0           0.0
AERTN BASIN        0.0         0.0          0.0         0.0           0.0
MECH AERATN        0.0         0.0          0.0         0.0           0.0
SECY SETTLER       0.0         0.0          0.0         0.0           0.0
RECIRC PMPNG       0.0         0.0          0.0         0.0           0.0
CHLORNE FEED       0.0         0.0          0.0         0.0           0.0
CHLORNE BASN       0.0         0.0          0.0         0.0           0.0
GRAV THICKNR      88.40       88.40        88.40        0.290         0.378
SLGE HLD TNK       0.0         0.0          0.0         0.0           0.0
VACUUM FLITR   17365.26    17365.30     17365.30       15.364        32.729
FECL FEEDING  299353.00   299353.06    299353.06       13.101       312.453
LIME FEEDING    9722.82     9722.85      9722.85        9.124        18.847

TOTALS =      326529.44   326529.56    326529.56       37.873       364.407
```

```
                         UNIT COST DATA
                       (CENTS/1000 GALLONS)

PROCESS NAME   AMORTIZATION   LABOR   MATERIAL & SUPPLY   TOTAL

RAW WSTE PMP       0.0         0.0          0.0             0.0
PRE TRETMENT       0.0         0.0          0.0             0.0
PRIM SETTLER       0.0         0.0          0.0             0.0
PRMYSLGE PMP       0.0         0.0          0.0             0.0
AERTN BASIN        0.0         0.0          0.0             0.0
MECH AERATN        0.0         0.0          0.0             0.0
SECY SETTLER       0.0         0.0          0.0             0.0
RECIRC PMPNG       0.0         0.0          0.0             0.0
CHLORNE FEED       0.0         0.0          0.0             0.0
CHLORNE BASN       0.0         0.0          0.0             0.0
GRAV THICKNR       0.04        0.00         0.00            0.05
SLGE HLD TNK       0.0         0.0          0.0             0.0
VACUUM FILTR       0.11        0.21         0.24            0.56
FECL FEEDING       0.13        0.18         4.10            4.41
LIME FEEDING       0.04        0.12         0.15            0.30

TOTALS =           0.32        0.52         4.47            5.31
                  SOLID SINKS COSTS

TIME      SINK   AMOUNT SOLIDS        COST
PERIOD    USED      (TONS)           ($/TP)
   1        1        2553.97       63849.36
   2        1        2553.97       63849.36
   3        1        2553.97       63849.36
   4        1        2553.97       63849.36
   5        1        2553.97       63849.36
   6        1        2553.97       63849.36
   7        1        2553.97       63849.36
   8        1        2553.97       63849.36
   9        1        2553.97       63849.36
  10        1        2553.97       63849.36
  11        1        2553.97       63849.36
  12        1        2553.97       63849.36
  13        1        2553.97       63849.36
  14        1        2553.97       63849.36
  15        1        2553.97       63849.36
  16        1        2553.97       63849.36
  17        1        2553.97       63849.36
  18        1        2553.97       63849.36
  19        1        2553.97       63849.36
  20        1        2553.97       63849.36

  TOTALS =          51079.45     1276985.00

UNIT COSTS FOR SOLID WASTE DISPOSAL (CENTS/1000  GAL.) = 0.87
TOTAL UNIT COST (CENTS/1000 GAL) = 6.18
```

APPENDIX B

METRIC CONVERSION CHART

Multiply	By	To Get
Inches	2.54	Centimeters
Feet	0.3048	Meters
Square Feet	0.0929	Square Meters
Cubic Feet	0.0283	Cubic Meters
Pounds	0.454	Kilograms
Gallons	3.79	Liters
Gallons/Minute	5.458	Cubic Meters/Day
Feet/Second	0.305	Meters/Second